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INCINERATOR OVERFIRE MIXING DEMONSTRATION

T. J. Lamb, et al

Arthur D. Little, Incorporated

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T. J. Lamb, R. H. Stephens, C. M. Mohr,
P. C. Levins, L. K. Fox, and A. F. Sarofim

Arthur D. Little, Inc.
20 Acorn Park
Cambridge, Massachusetts 02140

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EPA Project Officers:

James D. Kilgroe

Industrial Environmental Research Laboratory
Research Triangle Park, North Carolina 27711

and

Donald A. Oberacker

Municipal Environmental Research Laboratory
Cincinnati, Ohio 45268

Prepared for

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NOMENCLATURE

A	Area
A_c	Area of crossflow, Eqs. (5) and (14)
A_f	Area of flow above jets
A_j	Area of overfire jet
A_o	Refuse bed area
C	Concentration of off gas expressed as mol/volume or as mol/fraction
$\bar{C}$	Mean concentration
C_q	Concentration of off gas at onset of quenching
C_s	Concentration of off gas in stoichiometric mixture
C_T	Total molar concentration, mols/volume e.g. (P/RT)
C_p	Heat capacity
$C_{p_s}, C_{p_{xs}}$	Heat capacity of combustion products of stoichiometric mixture, of excess air
c'	Concentration fluctuations about the mean
$\sqrt{c'^2}$	Unmixedness factor (root mean square of c')
$(\sqrt{c'^2})_b, (\sqrt{c'^2})_{bo}$	Unmixedness factor for tests without jets with breech, over bed
$(\sqrt{c'^2})_j, (\sqrt{c'^2})_{jo}$	Unmixedness factor for tests with jets with breech, over bed
D_u	Diffusivity of fuel in air, gm/cm ² /sec, etc.
d_j	Diameter of jet
E_a, E_{af}, E_{ar}	Activation energy, for forward reaction, for reverse reaction
$f(\phi)$	$\sqrt{c'^2}/\Delta C$ defined by Figure 3
g	Acceleration of gravity
g_c	Conversion factor of weight to mass
HHV	Higher heating value
K	Equilibrium constant
K_{wg}	Equilibrium constant for water-gas shift
KE	Kinetic energy
k	Constant or thermal conductivity (App. C)
k_3	Chemical kinetic constant of reaction (3), for example
L	Characteristic dimension of system
L_p	Penetration distance of jet
m	Mass

NOMENCLATURE (Continued)

N	Number of jets
n	Moles per mole of off gas
n_{cp}	Moles of combustion products per mole of off gas
n_{o_2}	Moles of oxygen required to bar one mole of off gas
$n_{H_2O}, n_{N_2}, n_{CO_2}$	Moles of water vapor per mole of off gas, of nitrogen, of CO_2 , etc.
n_s	Total moles in stoichiometric mixture containing one mole of off gas
n_T	Total moles in mixture containing one mole of off gas
n_{xs}	Moles of excess air (not oxygen) per mole of off gas
$(O_2 \text{ Req})_b, (O_2 \text{ Req})_{b_o}$	Oxygen required to complete combustion of off gas without jets in breech, initial
$(O_2 \text{ Req})_j$	Oxygen required to complete combustion of off gas in breech with jets on
$(O_2 \text{ Req})_{j_o}$	Oxygen required to complete combustion of off gas at top of bed with jets on
P	Pressure
P_a	Partial pressure of component a, for example
$P_o - P$	Differential pressure between grate and breech
Q	Flow rate, volume/time
Q_c	Flow rate of cross flow to jet axis
Q_f	Flow rate of gases after jets or exit the breech
Q_j	Flow rate of jets
Q_o	Flow rate of off gas from refuse bed
R	Ideal gas constant
S	Jet spacing parameter
T	Temperature
T_a	Adiabatic flame temperature
T_C	Temperature of gas in cold zone of incinerator
T_H	Temperature of gas in hot zone of incinerator
T_o	Ambient temperature
T_j	Jet temperature
T_c	Temperature of crossflow gases
T_f	Temperature of gases after jets

NOMENCLATURE (Continued)

t	Time
u	Velocity
u_C, u_{C_0}	Cold zone velocity, initial value
u_c	Crossflow velocity (before jets)
u_f	Furnace velocity (after jets)
u_H, u_{H_0}	Hot zone velocity, initial value
u_j	Jet velocity
u_o	Initial velocity
V	Volume of gas
V_f	Furnace volume
x	Moles fraction of off gas in mixture of off gas plus jet air
y	Mole fraction of off gas
y_c	Mole fraction of off gas in crossflow (before jets) gases
y_f	Mole fraction of off gas in furnace (after jets) gases
y_s	Mole fraction of off gas in stoichiometric mixture
Δy_b	$(y - y_s)$ evaluated for base tests (no jets)
Δy_j	$(y - y_s)$ evaluated for jet tests
z	Distance
z_o	Height at top of refuse bed
z_C	Height to cold zone in breach
z_H	Height to hot zone in breach
z_q	Distance from initial surface to quench plane (App. C)
z_s	Distance from initial surface to stoichiometric plane (App. C)
z'	Distance from initial surface to zero-oxygen plane (App. C)

Greek Symbols

α	thermal diffusivity or constant [Eq. (F-4)]
β	mixing energy constant [Eq. (22)]
γ	ratio of furnace height to width
ϵ	rate of input of mixing energy per mass of system
η	ratio of furnace length to width, also constant [Eq. (C-12)]
ρ	density
ρ_C	density of cold zone gases
ρ_H	density of hot zone gases
ρ_c	density of crossflow gases
ρ_f	density of furnace gases (after jet)
ρ_j	density of jet air
τ	residence time
ϕ	available oxygen/oxygen required to complete combustion
ϕ_b, ϕ_{b_o}	ϕ for base tests (no jets), initial value at top of refuse bed
ϕ_j, ϕ_{j_o}	ϕ for jet tests, initial value at top of refuse bed

Common Subscripts

b	baseline (no jets)
C	cold zone
c	crossflow (before jets)
f	furnace (after jets)
H	hot zone
j	jet
o	initial, usually refers to top of refuse bed
q	quenched conditions
s	stoichiometric conditions

Gas Compositions

(% CO ₂)	percent CO ₂ in gas mixture
(% CO ₂) _b	percent CO ₂ without jets on
(% CO ₂) _j	percent CO ₂ with jets on
% CO ₂ _s	percent CO ₂ in stoichiometric mixture
CO	carbon monoxide (ppm)
CO _s	carbon monoxide in stoichiometric mixture with defined CO/CO ₂
% O ₂	percent oxygen

SECTION I

CONCLUSIONS

Overfired air jets have been used in coal stokers to induce turbulence for increased combustion efficiency and reduced emissions of combustible pollutants such as CO, H₂, hydrocarbons, and soot. A mixing system of this type was installed in an existing municipal incinerator under Contract 68-02-0204 with the Environmental Protection Agency. Tests were conducted to determine the effects of and design criteria for mixing jets. As a result of the testing program and subsequent analyses, Arthur D. Little, Inc., reached the following conclusions:

REDUCTION OF COMBUSTIBLE EMISSIONS

1. Combustible emissions can occur either by quenching where the temperature of the gases is too low to sustain combustion, or by a mixing limitation where the gases are not completely mixed with the available oxygen, and thus will not burn out in the furnace within the residence time. The former becomes significant when gas temperatures are below about 1400°F; the latter is significant when gas temperatures are above about 1800°F. By adjusting air input to maintain furnace temperature within the range of 1400 to 1800°F, both quenching and mixing limitation were avoided.
2. Overfire jets reduced combustible emissions resulting from the mixing limitation. This reduction is attributed to the greater availability of oxygen rather than to large increases in mixing intensity.

A theory based on statistical descriptions of the turbulent mixing was developed and verified by test data. This theory provides a quantitative expression for the heretofore qualitative concepts of the three T's — time, temperature, and turbulence. This theory can be used to show under what conditions combustible emissions caused by inadequate mixing can be reduced by increasing mixing intensity through the use of sidewall jets or by increasing furnace residence time.

3. Overfire jets were ineffective in reducing combustible emissions due to quenching. These emissions are believed to be more amenable to control through proper furnace design and underfire air control than through mixing. Unfortunately, the test incinerator was not sufficiently versatile to allow us to explore this emission cause in depth.

JET MIXING

1. Test results showed that the penetration of a sidewall jet into the gases from the refuse bed can be predicted using the design equations presented in this report.

2. Sidewall overfire jets were found to be very effective in controlling the temperature of the overfire region.
3. The jet test configurations, such as opposed jets, interlaced jets, or jets on one side, were shown to be equally effective for attaining adequate uniform mixing and temperature control.
4. While the jets did not have a significant effect on increasing the mixing intensity in the test furnace, extrapolations of the theory, which was verified in the test furnace, suggest that such an effect could be significant in smaller furnaces.

FURNACE FLUID FLOW

1. The characteristics of the stratified layers, including their acceleration due to buoyancy forces, can be predicted from the geometry of the furnaces and temperatures along the refuse bed surface.
2. When the gases exit directly over the cold end of the furnace (the burnout grate) so that no forced mixing of the hot gases and cold gases occurs, then the gases will stratify, forming stable layers. The colder gases are on the bottom and quench combustion, while the hotter gases are on top and are often oxygen-deficient. Both conditions can result in high combustible emissions.

BED BURNING

1. The processes of drying, ignition, pyrolysis, combustion, and burnout occur simultaneously throughout the refuse bed as witnessed by a constancy of the ratio H_2O/CO_2 throughout the furnace.
2. The composition of gases both leaving the refuse bed and also in oxygen-deficient (unmixed) pockets is controlled by the water/gas shift equilibrium, as confirmed by test data in both the breech of the test furnace and also directly over the refuse bed.
3. In a furnace that is unsealed, or which has a large amount of uncontrolled overfire air, the oxygen required for combustion is entrained into the refuse bed from the overfire air space. Even when the underfire air is considerably less than stoichiometric, the burning rate of refuse tends to be constant at approximately 60 lb/hr/ft^2 .

SECTION II

RECOMMENDATIONS

Based on the data collected in this experimental program, Arthur D. Little, Inc., recommends the following:

1. The use of overfire air jets to reduce combustible emissions from existing municipal incinerators.
2. The use of overfire jets in small furnaces to increase mixing intensity. Design criteria presented in this report should be utilized.
3. The realization of uniform mixing throughout the furnace through proper control of underfire air and certain changes in furnace configuration. For example:
 - a. To achieve uniform mixing with underfire air, a high-pressure drop grate should be used, and the furnace should be sealed to prevent any leakage which would tend to negate the effect of controlled air rates.
 - b. The furnace should be designed so that hot and cold gases are forced to mix before exiting the furnace. This represents the most effective technique for attaining a uniform front-to-back mixing.
 - c. Overfire jets should be designed to effect uniform side-to-side mixing, as shown in this report.

SECTION III

INTRODUCTION

BACKGROUND

The emissions from municipal incinerators have been the subject of considerable discussion over the past decade in conjunction with growing public awareness of the adverse effects of air pollution. The incinerator emission receiving the most attention to date is fly ash, a particulate emission having a characteristic dimension between 0.1 and 100 microns (μ). Approximately 70% (by weight) of fly ash has a characteristic dimension greater than 5μ and is therefore easily removed using relatively simple devices. The smaller particles are much more difficult to remove, however, but in recent years several advances have been made in the state-of-the-art of small particulate removal. The development of such devices as electrostatic precipitators and high-energy wet scrubbers — both high-efficiency collection devices — has facilitated the removal of particles down to 0.5μ so that 99% of the particulate emission can be controlled. As municipal incinerators across the country incorporate these types of emission controls, the prevalence of fly ash as a pollutant will diminish.

On the other hand, in its report [1] "Systems Study of Air Pollution from Municipal Incinerators" (Contract CPA-22-69-23 and dated March 1970), Arthur D. Little, Inc. (ADL) stated that the municipal incinerator was a large source of such pollutants as carbon monoxide, hydrocarbons and soot (small carbon particles). Unfortunately, these emissions are not so easily controlled as is the fly ash. The soot particles are so small, in fact, that a significant fraction will pass through even the highest efficiency control devices. The gaseous emissions are relatively unaffected by particulate control devices. However, these emissions are all products of incomplete combustion and, therefore, can be controlled by improving combustion efficiency rather than by the use of expensive air pollution control devices. In its report ADL recommended that methods be developed to improve the combustion efficiency of municipal incinerators to preclude incomplete combustion.

Designers have long recognized that combustible emissions could be reduced by increasing the combustion efficiency of a process. In fact, they have taken several steps to maximize residence time and increase the amount of mixing within the furnace, but most current design practice is based upon empirical rules or qualitative guidelines. Unfortunately, many accepted engineering principles of combustion have not as yet been applied to incinerators.

In February 1972, ADL published another report entitled, "Incinerator Overfire Mixing Study" (Contract EHSD 71-6) [2]. This study, which reviewed the present understanding of the combustion process in solid fuel equipment and municipal incinerators, postulated a technique for improving combustion efficiency in which jets would be used to promote mixing. A test program was proposed to demonstrate the use of overfire air and steam jets to reduce the combustible pollutants being emitted from an existing municipal incinerator.

In this previous study, ADL postulated that combustion in a municipal incinerator occurred in three distinct zones – an ignition-limited zone, a combustion-limited zone, and a fuel-limited (burnout) zone. Under conditions normally found in incinerators, the oxygen concentrations in the first and last zones are high, reflecting the low oxygen demand required to sustain combustion. The central zone is oxygen-limited and has an extremely low residual oxygen concentration. Similarly, the temperature in the central zone is considerably higher than in either of the other two zones and the gas emanating from this zone has a much greater concentration of CO and other combustible emissions. Unfortunately, the oxygen-rich zones do not mix well with the oxygen-deficient central zone. The model developed to determine gas-flow patterns within the incinerator indicated the zones remain stratified long after the combustion process had been quenched. We, therefore, suggested two approaches to eliminate the emissions resulting from incomplete combustion:

1. Distribution of underfire air along the grates in proportion to the oxygen demand, and
2. The use of air and/or steam jets to induce turbulence, thus promoting greater combustion efficiency.

To verify the model for burning, mixing, and stratification and to demonstrate the effectiveness of overfire mixing to reduce combustible emissions, EPA awarded a contract to ADL to test an existing municipal incinerator and to develop design and operating guidelines which could be used for improving the combustion efficiency of other incinerators.

OBJECTIVES

Compared to other industrial sources of pollutants, the municipal incinerator is a low-budget piece of equipment, often poorly operated or maintained. There has been little incentive for either sophisticated testing or development work on the part of either the municipality (the owner) or the engineer (designer). For that reason very few meaningful data are available.

To extend the state-of-the-art, this program had several independent objectives. They were to:

1. Develop a basic understanding of the incineration process by verifying the three models presented in the previous report and extending these concepts as far as possible to describe completely the combustion of refuse within an incinerator;
2. Demonstrate the use of overfire jets on an existing municipal incinerator of common design and showing their effect on combustible emissions;

3. Develop guidelines for design and operation of the jet systems, because much of the work is of a theoretical nature extending beyond the normal interests of both the designer and the operator; and
4. Supply these data insofar as possible, because — independent of the focus of the other objectives — there is an extreme lack of basic operating and emission data on incinerators.

APPROACH AND SCOPE

In developing design data for the use of overfire jets on municipal incinerators, we tried to strike a balance between both the theoretical and the experimental approaches that are required to satisfy the specific objectives of this program. Although we only tested a single incinerator, our intent was to generalize the test results to other furnaces.

Full characterization of the incinerator would involve a complete input/output analysis. Since the principal emphasis in the program was on combustible emissions, the major data-collection effort was restricted to this area. Limited data on material input were obtained to guide the overall data analysis. The information obtained during each of the tests included:

- Input: refuse tonnage, grate speeds and air flow rate;
- Output: temperature, flow (velocity), and composition.

Gaseous composition data were required to reflect both the overall combustion performance of the incinerator and the variation in combustible emissions. The following species were routinely analyzed in the breaching region (and over the burning bed during the study of that region):

Overall composition: O_2 , CO_2 , N_2 , and H_2O ;

Combustible emissions: CO , H_2 , and THC (total gaseous hydrocarbons).

In addition, a carbon monoxide analyzer was run continuously in the duct leading to the base of the stack to reflect the long-term trend in changes of emissions.

The key factor in the analytical testing program was the development of a high-temperature testing probe to sample in the $2000^\circ F$ (or greater) conditions found both in the breach and over the burning bed. Continued service from a test probe under these conditions for more than one week of testing proved to be very difficult, but because of modification in the testing procedure and improvements in the probe design, we believe that the technique developed in this program will be generally applicable to all similar instances of emission testing in high-temperature environments.

In conducting this program, the initial series of tests (the baseline tests) was used primarily to verify the analytical approaches and to ensure that the test data were consistent

internally and with the physical concepts of incineration. With each series of tests, material and energy balances were completed and the results were analyzed to indicate potential weaknesses in the test data or to point out areas where present understanding of combustion could not correlate the experimental results. Because of this procedure, there were changes made in the analytical approach with corresponding increases in the accuracy of the experimental data. The material balances for the final jet tests indicated the magnitude of error in the analytical data to be less than 10%.

Upon verification of the initial testing data, it became apparent that many of the theories used to model the combustion were not wholly adequate with respect to the characterization of mixing and combustible emission burnout. Additional work in the modeling of combustion was carried out to augment the models presented in our previous report. The emphasis was on increasing the sophistication of the models so that the physical phenomena could be more readily deduced from the test data. The analysis of the bed model took on much less importance as the work progressed. The bed model is introduced only in the Summary to the report. The detailed discussion of the bed model is contained in Appendix E.

We did not intend for this report to be an exhaustive compendium of combustion mechanics, but only that the principles and experimental results pertinent to the burnout of combustible emission be fully explored and documented. We have incorporated the test data in as complete a form as possible (Appendix A), and we feel it will prove to be a valuable resource to those workers in the field who would like to extend this work still further, or who may require data on existing furnaces for experimental support of their own theories.

SECTION IV

SUMMARY

The primary objective of this program was the reduction of combustible emissions from existing municipal incinerators through the use of overfire air jets to induce mixing. The objectives cited in the Introduction of this report represent the various aspects required to study this particular problem adequately.

TESTING PROGRAM

To provide the data required to verify the burning and mixing models developed in the Incinerator Overfire Mixing Study (2), we planned a test program at an existing municipal incinerator. We chose the incinerator of the City of Newton, Massachusetts, as the test vehicle, because it is similar in design to many incinerators in the United States, it could be easily retrofitted with an air jet system, and the City of Newton was kind enough to make the facility available.

In the initial tests, we recorded composition, temperature, and velocity data in the breech of the furnace to check out the analytical procedures and provide information needed to design the jet system.

Following the procedure outlined in the previous study, we designed and installed an overfire jet mixing system. The system consisted of ten 4-inch jets on each side of the furnace, a forced draft fan, and the ducting and instrumentation necessary to control the system.

The test program included tests both in the breech of the incinerator and also over the refuse bed. The breech tests were required to verify the fluid flow model and also to confirm or deny the effectiveness of overfire mixing jets in reducing combustible emissions. The bed data were necessary for verifying the bed burning model.

VERIFICATION OF MODELS

In the previous study (2), ADL postulated three models necessary to describe incineration: 1) a bed-burning model, 2) a fluid-flow model, and 3) a jet-mixing model. All three were verified either as postulated or with the refinements added as a result of the testing work. Each model is summarized below.

Bed-Burning Model

The bed-burning model was originally proposed to describe the overbed gas composition and also the burning rates for refuse beds. The model was previously postulated on the basis of separate drying, ignition, pyrolysis, combustion and burnout planes, but the results

of the testing program showed that the processes occurred simultaneously throughout the refuse beds. However, this fact does not alter the concept that the gases leaving the refuse bed in the absence of oxygen are controlled by the water gas shift equilibrium. In support of this, gas samples taken directly over the refuse bed contained concentrations of CO, CO₂, H₂, and H₂O in direct proportion to what would have been predicted from the equilibrium relation. However, each sample also contained a significant amount of oxygen, although the previous model was based upon the existence of a plane of zero oxygen within the refuse bed. This particular phenomenon strongly suggested that the combustible gases were not mixed with oxygen and remained unmixed even after the gases had left the furnace. As a result a fourth model – the microscale mixing theory – was deemed necessary to describe completely the combustion phenomenon that was occurring within the furnace volume. This model is described in a later section of this summary and also in Section V.

Based upon the empirically determined amounts of oxygen that can be entrained in the refuse bed and also upon the water gas shift gas equilibrium, we were able to develop estimates for overbed composition of gases. This information was required in the model developed for microscale combustion.

An additional ramification of the oxygen entrainment was the fact that the bed-burning rates tended to remain constant at approximately 60 lb/hr/ft². Surprisingly enough, the bed-burning rate was not very sensitive to the underfire air rate in that the air deficiency was supplied from the overfire region. This would not be the case either when the amount of oxygen was very limited, or when greater than stoichiometric amounts of underfire air were present.

Fluid Flow Model

The acceleration of gases away from the refuse bed can be mathematically characterized using Bernoulli's equation:

$$u_c^2 = u_o^2 + 2g(z - z_o) \left[\frac{T_h}{T_c} - 1 \right]. \quad (1)$$

In a furnace that is completely sealed, the appropriate T_c is the temperature in the cold region of the furnace, most likely the temperature over the burnout grate. In an unsealed furnace the appropriate T_c is the ambient air temperature. Because of the large amount of leakage that was inherent in all of the test runs at the Newton incinerator, it was obvious that the furnace was unsealed. Equation (1) was verified for a T_c of approximately 70°F.

Jet Mixing Model

In a previous study (2), ADL postulated that jets could be used to mix cold regions, such as the region over the burnout grate, with the hotter regions of the furnace so that the overall temperature of the furnace could be controlled. To design such a jet it was necessary

to arrive at a relationship between the jet penetration distance and the parameters of the system. The jet tests were able to verify the design equation suggested by Ivanov (3):

$$L_p = 1.6 \frac{u_j d_j}{u_c} \sqrt{\frac{\rho_j}{\rho_c}} \quad (2)$$

In designing a jet system, the designer has the freedom to adjust the jet diameter location and the jet velocity, with the only constraint being that the jet system not be allowed to deliver an amount of air to the furnace that would quench combustion. The other variable in the jet design equation is u_c , the cross-flow velocity for the gases rising off the refuse bed perpendicular to the row of the jets. This is determined from Eq.(1) above.

STRATEGY FOR EMISSION CONTROL

Origin of Combustible Emissions

Two phenomena which result in combustible emissions were identified in the course of this study. The experimentally determined CO emission factor for the furnace is a function of local furnace temperature as shown in Figure 1. In regions of the furnace where the temperature was below approximately 1100°F, emissions occurred because the reaction had been quenched, i.e., it was too cold to sustain combustion.

On the other hand, when the temperature was high (above 1800°F), the corresponding amount of excess air available was very low. In the case of perfect mixing, this has no effect, but for incineration, we postulated that the gases leaving the furnace have a heterogeneous composition consisting of complete separate pockets of oxygen containing combustion products and "fuel" containing combustion products. At high temperatures "fuel" and air do not co-exist in a pocket for long, because the combustion reactions occur too rapidly. When the amount of available oxygen decreases (as temperature increases), the probability of finding a pocket of "fuel" that has not mixed with available oxygen increases. This mixing limitation, as indicated by higher CO emissions, was observed for temperatures above about 1800°F.

Reduction in Emissions

The key to the elimination of combustible emissions is enhancement of combustion efficiency in the overfire region of the furnace. For a long time designers have recognized the importance of the three T's – time, temperature, and turbulence. In this report we have gone one step farther in developing the necessary mathematical models for describing the various phases of the operation so that the interrelationships between the three T's can be quantitatively described.

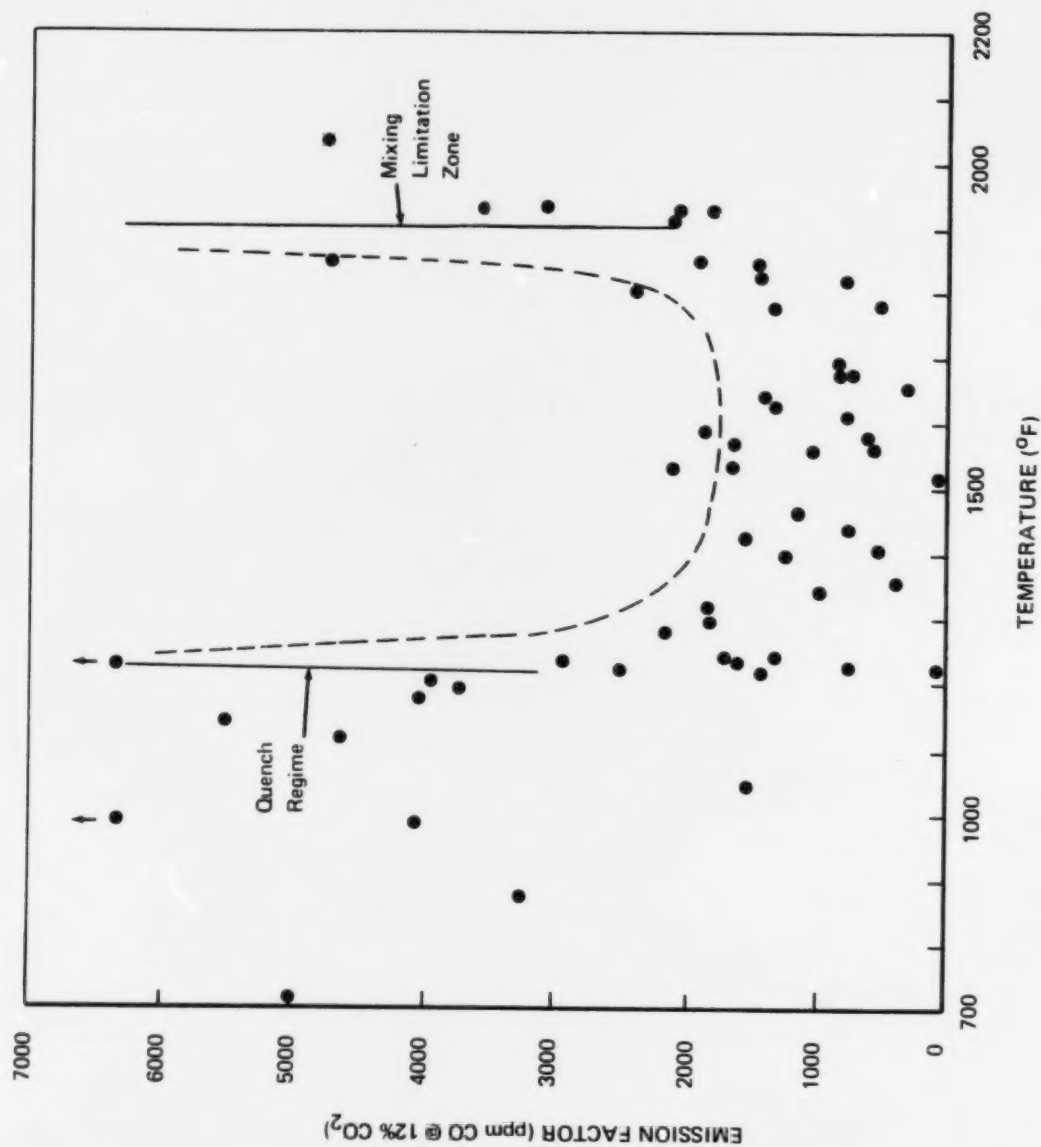


Figure 1 Emission factors vs. temperature.

TEMPERATURE CONTROL

The data of Figure 1 clearly show the higher CO emission at temperatures above about 1800°F or below about 1400°F. We believe the high-temperature emissions are caused by a mixing limitation resulting from a relatively low amount of available oxygen (excess air). As more excess air is added (consequently lowering gas temperature), the mixing constraint is eased and the high CO emissions are reduced. In the case of high-temperature emissions, jets can be used to add a sufficient amount of overfire air to control the excess air within the given range. In designing such a system, it is important to obtain a uniform "macroscale" mixing throughout the furnace. The specific design criteria are discussed later in this section.

Using this type of control, the low levels of combustible emissions (as obtained in Figure 1) can be achieved.

Mixing

As pointed out above, it is important to maintain a uniform macroscale mixing throughout the furnace, primarily to avoid hot and cold spots, i.e., to maintain the gas within the 1400 to 1800°F range. However, in terms of burnout of combustible emissions one must also consider the problem of microscale mixing in addition to the macroscale considerations of uniformity. The microscale problem is concerned with the mixing within a local region of the furnace, or along a given streamline, to bring the combustibles and available oxygen into intimate contact with each other so that the desired burnout can occur.

To describe this macroscale mixing, it was necessary to introduce two well known concepts that have not been previously applied to incinerator designs. These are:

- (1) The use of the unmixedness factor to characterize the level of emissions; and
- (2) The use of statistically based mixing theories to describe the relationship between the degree of mixing and the time required to burn out a given pocket of combustibles.

These concepts are discussed in detail in Section V. A qualitative discussion of these concepts will be included here.

- Unmixedness

For complete combustion, mixing on a molecular level is required. A single sample containing both oxygen and combustibles reflects alternate sampling from oxygen-rich and fuel-rich gas pockets. A measure of unmixedness of the overall sample can be deduced from the ratio of the concentration of oxygen available in the gas sample to that required to complete the combustion process using the statistical description of turbulence. This measure [as

suggested by Hottel (4)] reflects both the intensity of turbulence as well as the relative concentration of the combustibles and air.

The emission data shown in Figure 1 correlate with furnace temperature and Hottel's (4) unmixedness factor to the extent that the scatter in the data can be resolved within about 10%.

- Microscale Turbulent Mixing

The final intimate mixing of oxygen and fuel can be described as a turbulent fluctuation decay process analogous to eddy decay phenomena that have been extensively studied in wind tunnel experiments.

The mathematical relation suggested by Corrsin (5) equates the unmixedness factor ($\sqrt{c'^2}$) to an exponential function of residence time and mixing intensity (turbulence).

Mixing intensity and residence time are both affected by the furnace design. Since the unmixedness factor can be related to design parameters, based on Corrsin's equation, and to the emission factor, based upon Hottel's relation, this defines the overall quantitative relation between emissions and the three T's. The data taken in this program were sufficiently accurate to support this conclusion, but not sufficiently accurate to provide proof of the hypotheses.

DESIGN GUIDELINES

To achieve the desired reduction in combustible emissions, a proper design must address both the macroscale and microscale mixing effects. The uniform, or macroscale, mixing can be achieved both through the proper use of overfire jets and through careful design of the furnace enclosure, considering both the configuration and furnace dimensions. The jet design strategy is based upon the method of Ivanov (3) which could be easily verified from the test data. Several tests were run using various jet positions, including interlaced, opposed, and one-sided jets, and all were shown to be equally effective in providing uniform macroscopic mixing and temperature control capabilities. The jets proved to be most advantageous when used to achieve an overall side-to-side mixing and to maintain the overfire temperature within the proper range.

Overfire jets have been suggested also for use as a means of obtaining front-to-back mixing in the furnace. However, we believe the same effect would be more easily obtained through the proper design of the furnace configuration and proper control of the air distribution so that quenching zones are prevented from occurring, and so that hotter gases are mixed with the colder gases to attain a uniform temperature throughout the furnace. The latter is achieved by designing a furnace with flow patterns so the hot gases are deflected into the colder gases, or in some other way forced to mix with the colder gases, rather than allowing stratification to occur, as that demonstrated in the testing program.

In addition, the furnace should be a sealed enclosure so that maximum control of the air is maintained through the elimination of leakage. Moreover, the roof of the furnace should be fairly close to the refuse bed so that the gases rising from the bed do not have a chance to accelerate under the influence of buoyancy forces. By limiting the velocity of the gases, the jets will be able to achieve a much greater control of the overall mixing patterns within the furnace.

Microscale Effects

Overfire jets can also be used to increase the mixing intensity of the overfire gases, but care must be taken to avoid consequential quenching of the combustion reactions. The primary use of the jets is for the macroscale mixing and temperature control, but if the jets can be so designed to accomplish these major objectives and still be capable of adding sufficient mixing power to the overfire regions to increase the microscale turbulence level, then so much the better. In our experience during the tests, we found that the amount of power that would be required to attain a noticeable effect on mixing intensity was extremely large so as to be prohibitive. The mixing power required to achieve a given mixing intensity scales as the fifth power of the characteristic dimension of the furnace. Calculations show that for the large municipal incinerator the effects of jets on microscale mixing are only marginal, being effective in some cases and ineffective in other cases. But in smaller furnaces, the power requirements are greatly reduced so that jets could be significant in affecting the mixing intensities. As a result of the tests, we were able to develop a design criterion to determine in which cases the jets could be used to attain a mixing intensity advantage, in addition to accomplishing the macroscale mixing and temperature control functions. This is discussed in Section VIII. The criterion is based upon the quantitative relationship between the emission factor and the three T's that was developed during the course of the program. The full utilization of this criterion and of the other conclusions derived from this program will represent a major step forward in the development of incineration technology within the framework of combustion engineering.

SECTION V

CONCEPTS IN INCINERATION

Incineration has proven very difficult to describe quantitatively, because of the simultaneous occurrence of a great number of very complex phenomena. The description of a combustion process must include fluid mechanics, thermodynamics, heat and mass transfer, reaction kinetics, and statistics. However, these theoretical principles for the most part have been reduced to rules-of-thumb when applied to the design of an incinerator. For example, designers assume burning rates of 60 lb/hr-ft^2 and heat release rates of $25,000 \text{ Btu's/hr-ft}^3$, basing their designs on past experience or upon attempts to maximize the effects of the three T's: time, temperature, and turbulence.

In this chapter, we will describe the processes which occur during incineration, using the mathematical models developed as a part of this and the previous study on overfire mixing (2). In addition, we will propose a fundamental basis for many of the common design rules. Required background material related to the basic principles of combustion — chemistry, kinetics, equilibrium, and the like — is given in Appendix D.

STRATEGY FOR EMISSION CONTROL

Two phenomena which result in combustible emissions were identified in the course of this study. The experimentally determined emission factor for the furnace was shown as a function of furnace temperature in Figure 1. In regions of the furnace where the temperature was below approximately 1100°F , emissions occurred because the reaction had been quenched, i.e., it was too cold to sustain combustion.

High levels of emissions are also associated with high furnace temperatures. It is not uncommon to measure concentrations of CO or H_2 on the order of 1000 ppm in flue gases containing 10% O_2 and at 2000°F . These emissions are the result of incomplete mixing, i.e., combustibles and oxygen have not come into intimate contact at any time during the entire passage through the furnace.

The key to the reduction of both sources of combustibles emission lies in proper control of residence time, furnace temperature, and turbulence. Residence time is a design parameter of the furnace. During the testing of the existing incinerator this parameter could not be varied. The dramatic effect of local furnace temperature can be seen in Figure 1. The third parameter — turbulence — and the related rate of mixing are more difficult to characterize than either time or temperature. However, mixing is much more controllable in tests than the other two parameters. For example, the most direct approach to controlling high-temperature emissions is to add more air; this process is controlled by mixing. As another example, a possible approach to eliminating cold-temperature emissions is to force gases emanating from relatively cold areas of the furnace to pass through the hotter gases in the furnace. This process is also controlled by mixing.

EFFECT OF MIXING

In the test program, high-velocity overfire air jets were installed on an existing incinerator so that the effects of overfire mixing could be more fully explored. During the course of the work two mixing concepts emerged:

- 1) *Macroscale mixing* – required to obtain uniformity of gases throughout the furnace so as to eliminate or minimize the effects of cold or hot spots; and
- 2) *Microscale mixing* – required to bring molecules of oxygen and combustible gases together so that complete burnout of the combustible emissions could be obtained.

These two scales of mixing are completely different in character and are discussed separately. However, the optimum strategy for reduction of combustibles is common to both mixing scales, i.e., supply the available oxygen at selected locations in sufficient excess to burn out the combustibles in a reasonable time without quenching the reactions.

Macroscale Mixing

Non-uniform mixing throughout the furnace is quite common because of the extremely diverse combustion conditions observed over individual grates. Proper use of air jets firing across the fuel bed results in two benefits on the macroscale:

- 1) The jets supply additional oxygen so that the stoichiometry can be adjusted to the optimum level for control of high-temperature emissions; and
- 2) Jets can break up the stratification of the combustion gases so that oxygen-deficient (hot) regions and oxygen-rich (cold) regions are mixed.

To perform the required mixing functions, a properly designed jet must:

- penetrate far enough into the combustion chamber to influence the macroscale flow patterns; and
- carry sufficient oxygen to provide for combustion of the fuel components present in the gas, but not so much as to quench the reaction mixture.

Careful design is required to satisfy both criteria.

Jet Penetration – The design procedure for an overfire jet system was worked out in the previous mixing study based on the method of Ivanov [3]. In summary, an appropriately designed jet is one in which the momentum forces predominate over buoyancy forces. Using these criteria and considering the basic fluid mechanics of turbulent jets, Ivanov presented an equation for the penetration of the jet as follows:

$$L_p = 1.6 \sqrt{\frac{\rho_j}{\rho_c}} \frac{d_j u_j}{u_c} \quad (3)$$

where

- L_p = penetration distance
- ρ = gas density
- d_j = jet diameter
- u = velocity

and subscripts

- j = jet
- c = crossflow (in furnace).

Based upon an empirical analysis of jet interactions, Ivanov suggested a jet spacing of between 3 and 5 jet diameters. Since the flow rate from a given jet can be calculated from the equation:

$$Q_j = \frac{\pi d_j^2}{4} \cdot u_j = \frac{\pi}{4} d_j \frac{L_p u_c}{1.6} \sqrt{\frac{\rho_c}{\rho_j}} \quad (4)$$

and each jet covers an area of grate:

$$A_c = (S d_j) L_p \quad (5)$$

where

S = the spacing parameter (between 3 and 5),

then the amount of air introduced over a unit area of grate is:

$$q_j = \frac{Q_j}{A_c} = \frac{\pi}{4} \frac{u_c}{1.6 S} \cdot \sqrt{\frac{\rho_c}{\rho_j}} \quad (6)$$

which is dependent upon only the crossflow velocity, the spacing parameter, and the density ratio. The crossflow velocity can be determined from the model for furnace flow as discussed below.

Furnace Flow – In most existing municipal incinerators, flames are concentrated in the initial portions of the furnace and are considerably diminished toward the end of the grates where the refuse burns out. Previous models of the fluid-flow patterns in an incinerator have postulated the existence of stratified zones [2]. Hot furnace gases from the front end of the

furnace tend to rise because of buoyancy and exit the furnaces along the roof of the breech. Colder gases from the after sections form a layer exiting along the bottom of the breech. The two zones are stable and will not mix. The colder zone is often below the quench temperature, resulting in high levels of combustibles. Provisions for mixing of the two stratified zones will create a uniform condition more conducive to combustible burnout.

In our previous mixing study [2], we used a two-zone model to characterize the differences between the hot gases and cold gases in the furnace. Bernoulli's equation can be applied to both the hot zone and the cold zone, and the two equations can be combined as follows:

$$\begin{array}{c} \text{Hot Zone} \\ \hline u_H^2 = u_{H0}^2 + 2 \frac{(P_0 - P)}{\rho_H} g_c - 2g(z - z_0) \end{array} \quad \begin{array}{c} \text{Cold Zone} \\ \hline \frac{(P_0 - P)}{\rho_c} \frac{g_c}{g} = (z - z_0) \end{array} \quad (7)$$

Combining ($\Delta P_H = \Delta P_c$) yields:

$$u_H^2 = u_{H0}^2 + 2g(z - z_0) \left(\frac{T_H}{T_C} - 1 \right) \quad u_C^2 = u_{C0}^2 \quad (8)$$

The hot gases accelerate in traversing from the bed to the breech as a result of the net buoyancy force created by the temperature difference between the hot and cold zones. As the temperature of the hot zone approaches that of the cold zone, the amount of acceleration is reduced. Note that the cold gases do not accelerate, but leave the furnace at a constant velocity. If the furnace is not sealed (as in the case of the Newton incinerator), then there is an interaction between the inside and outside of the furnace such that both the hot and cold zones accelerate with the hot zone accelerating faster than the cold zone. The equations for flow in this case are based on the ambient temperature T_0 as follows:

$$\begin{array}{c} \text{Hot Zone} \\ \hline u_H^2 = u_{H0}^2 + 2g(z_H - z_0) \left[\frac{T_H}{T_0} - 1 \right] \end{array} \quad \begin{array}{c} \text{Cold Zone} \\ \hline u_C^2 = u_{C0}^2 + 2g(z_C - z_0) \left[\frac{T_C}{T_0} - 1 \right] \end{array} \quad (9)$$

Combining the two equations and assuming that $(z_H - z_0)$ equals $(z_C - z_0)$ yields:

$$u_H^2 - u_C^2 = (u_{H0}^2 - u_{C0}^2) + 2g \left[(z - z_0) \left(\frac{T_H - T_C}{T_0} \right) \right] \quad (10)$$

To digress for the moment, note that the time required for the gas to travel a distance ℓ along a streamline can be calculated by integrating the equation about the streamline:

$$\tau = \int_0^\ell d\ell(x, z) / u_z \quad (11)$$

The residence time can be calculated from this relation if the path of the gas is known. An analytical solution to this problem is quite complex and graphical procedures are therefore preferred. In determining the time required to reach a given position over the bed, a simplifying assumption that the direction of motion near the bed is always vertical results in an analytical solution to the above equation for the hot zone:

$$\tau = \frac{T_0}{g(T_H - T_0)} \left\{ \left[u_{H0}^2 + 2g \left(\frac{T_H - T_0}{T_0} \right) (z - z_0) \right]^{1/2} - u_{H0} \right\} \quad (12)$$

Note that if L'Hospital's rule is applied in taking the limit as T_H approaches T_0 , the equation reduces to:

$$\tau = (z - z_0)/u_0 \quad (13)$$

This is a form of the familiar equation:

$$\text{Time} = \text{Distance/Rate}$$

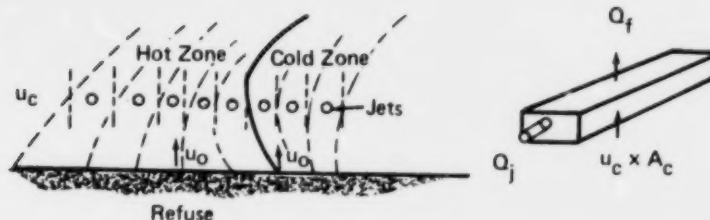
which applies to the cold zone (or any non-accelerating zone) under the assumptions made.

Stoichiometry – The air rate required to achieve the optimum stoichiometry is not at all a function of fluid mechanics, but rather is dependent upon a difference between the actual and the optimum concentrations of offgas emanating from the bed. In the section on microscale mixing, methods will be discussed for estimating these optimum concentrations. Consider the effect of the jet in mixing air with the offgas from the bed. A material balance can be made around the zone of jet action as shown in the illustration below. A material balance, assuming no change in temperature in going from the bed to the jet level, results in the equation:

$$u_0 A_0 = u_c A_c \quad (14)$$

where

- u_c = crossflow velocity,
- u_0 = initial velocity leaving the bed,
- A_c = crossflow area at the jet level, and
- A_0 = bed area.



Since the area of coverage of each jet is fixed by the jet design, according to Eq. (5) above, then the amount of gas coming from the bed that must be stoichiometrically adjusted by a given jet can be calculated. Note that if the jets are evenly spaced, then the jets in the hotter sections of the furnace interact with gases from proportionately larger grate areas than the jets in the colder section of the furnace because of the "necking down" process, common with accelerating gas streams. The volumetric flow into the zone of influence of the jet is therefore equal to $u_c \times A_c$. On the basis of ideal gases, a mass balance can be made about each jet zone as shown in Eq. (15) below:

$$\frac{Q_j|_{T_j}}{T_j} + \frac{Q_c|_{T_c}}{T_c} = \frac{Q_f|_{T_f}}{T_f} \quad (15)$$

where $Q|_T$ = volumetric flow rate evaluated at temperature T .

and subscripts j , c , and f = jet, cross-flow (before jets), and furnace (after jets), respectively.

Finally, a material balance of the offgas leads to the relation:

$$y_c \cdot C_{T_c} \cdot Q_c|_{T_c} = y_f \cdot C_{T_f} \cdot Q_f|_{T_f} \quad (16)$$

where y = mol fraction of "offgas," and
 C_T = molar concentration ($C_T = P/RT$).

and the subscripts c = crossflow condition (before jets), f = furnace condition (after the jets).

This equation can be combined with Eq. (15) and the ideal gas law to derive the relation:

$$\frac{y_c - y_f}{y_f} = \frac{Q_j|_{T_j}}{Q_c|_{T_c}} \left(\frac{T_c}{T_j} \right) \quad (17)$$

By rearranging and substituting for Q_c the jet flow rate can be determined by:

$$Q_j|_{T_j} = \left(\frac{y_c - y_f}{y_f} \right) \left(\frac{T_j}{T_c} \right) (S d_j L_p u_c) \quad (18)$$

This equation differs from the one based on jet penetration criterion Eq. (4), not in the design variables d_j , L_p , and u_c , but in the spacing parameter S , the temperature ratio, and the concentration terms. If the concentration term is determined to achieve the optimum emission level, i.e., if C_f is the concentration at the minimum point of the U-shaped emissions curve (Figure 1), then there is no assurance that complete penetration will be achieved. Likewise, if the jet flow rate is designed for complete penetration, there is no assurance that the optimum stoichiometry above the bed will be attained. The skill of the designer is then important in being able to balance the tradeoffs between complete penetration and optimum stoichiometry.

To further specify the design, note that Eqs. (4) and (18) could be set equal to each other and simplified to the relation:

$$\left(\frac{y_c - y_f}{y_f} \right) = \frac{\pi}{6.4S} \sqrt{T_c/T_j} \quad (19)$$

If the composition of offgas emanating from the bed could be predicted, then the left side of Eq. (19) could be evaluated. The equation would then provide a basis for determining S or the number of banks of jets that could be used to introduce all of the required air. Alternatively, the jet penetration criteria could be used to determine the amount of overfire air required. Equation (19) then forms the basis for determining the underfire air distribution necessary to produce the optimum stoichiometry. At the present time, our level of understanding does not allow us this additional degree of freedom.

Microscale Mixing

The co-existence of combustibles such as CO or H₂ and oxygen at high temperatures suggests that the gas has a heterogeneous consistency, as shown in Figure 2. The rate of dissipation and eventual burnout of the entrained pockets of combustibles is controlled by mixing at the microscopic level, i.e., microscale mixing. The problem is to correlate this mixing mechanism with the reduction of emissions.

The characterization here is adapted from turbulent fluid mechanics and mixing studies in high-velocity burners. Two distinct concepts are required to complete the characterization:

- 1) The relation of emission factors to fluid mechanics concepts, and

- 2) The relation of fluid mechanics concepts to incinerator design parameters (the three T's).

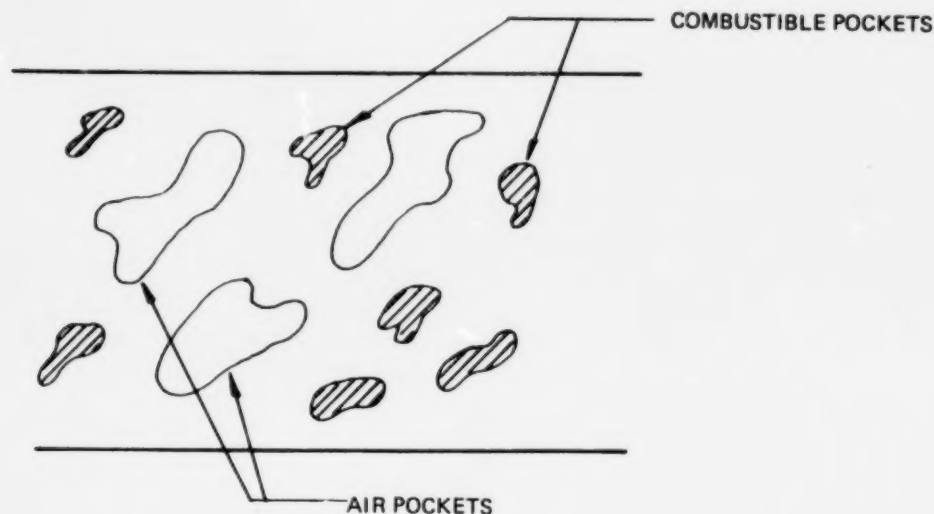


Figure 2 Heterogeneous character of flue gases

Emission Factors and Fluid Mechanics – The mixture of complete and partial combustion products emanating from the refuse bed is a gaseous "fuel," and the mixing process can be characterized as being similar to the mixing in turbulent flames (as suggested by Hawthorne, Wendell, and Hottel [4]). The characterization which results in a statistical description of turbulence can be summarized as follows.

In turbulent systems, concentration at any point in the system can be represented by two components: (1) the time mean value concentration, and (2) the fluctuating component of the concentration. These are represented by the equation:

$$C = \bar{C} + c' \quad (20)$$

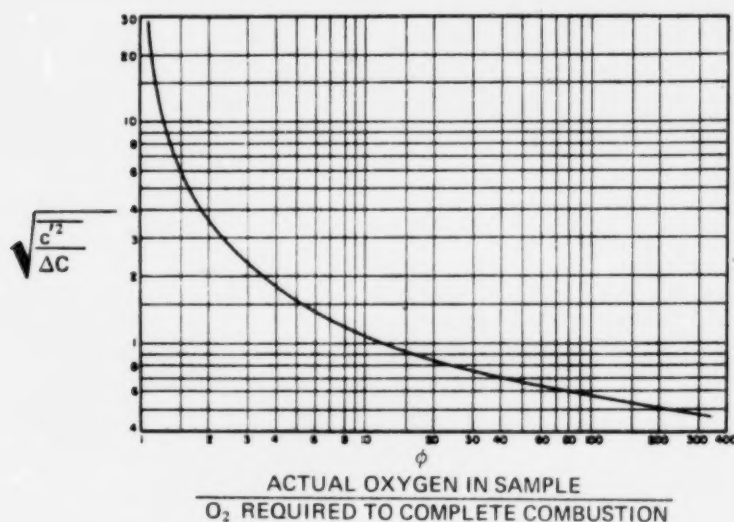
where c' has been shown to be randomly distributed about $\bar{C}$. The units of C can be either moles/volume or mole fractions.

By applying the gaussian distribution to c' , Hawthorne et al (4) showed that the ratio ϕ , of oxygen in a sample to the oxygen required to complete the combustion of the sample could be used to determine the values of the parameter $\sqrt{c'^2}/(C - C_s)^*$ where $\sqrt{c'^2}$ is the r.m.s. value of c' and $(C - C_s)$ is the difference between the initial "fuel" concentration C before any combustion occurs and the stoichiometric concentration C_s , i.e., that initial concentration of "fuel" which would just consume all oxygen present. Note that the variable C represents the concentration of "offgas" in air and not the concentration of the

*Hawthorne's derivation uses C to represent either mol fraction or molar concentration interchangeably. The same convention is used throughout this report where C is often used interchangeably with y , the mol fraction of offgas.

constituents making up the offgas. Also, note that C is not an average value of the concentration like C_s , but rather an instantaneous value which is a function of time and position.

Hawthorne et al (4) further showed that ϕ which can be calculated from the sample analysis, can be used to determine the nature of the unmixedness factor $\sqrt{c'^2}$, based upon the relation shown in Figure 3. The derivation of this relation is in Appendix F. An example of this calculation is given in Appendix B. The relation will be shown later to correlate quite well with the emission data for incinerators.



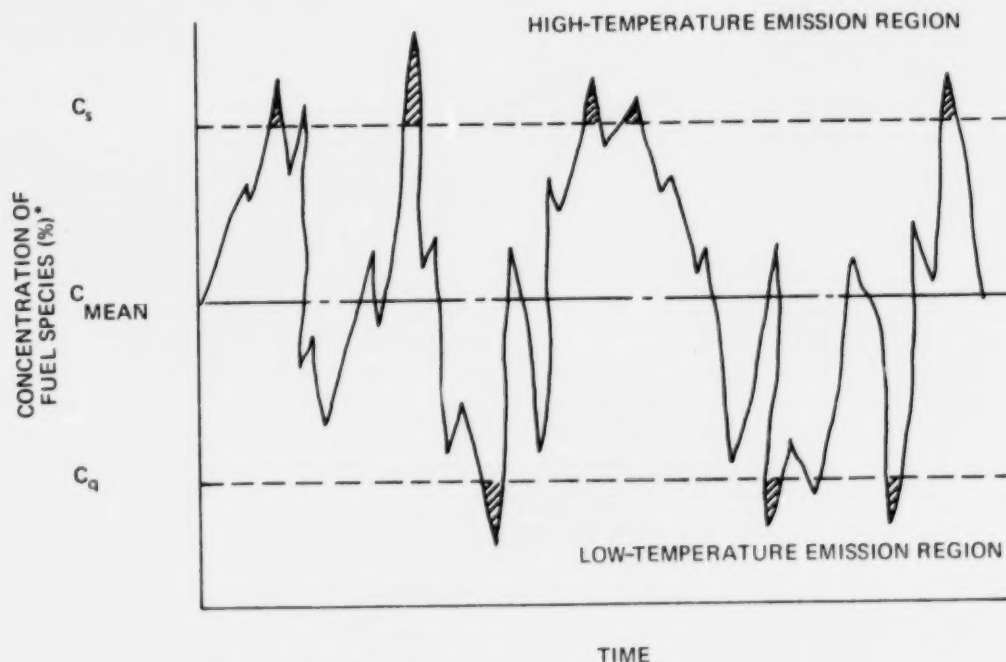
Source: Ref. (4).

Figure 3 The unmixedness factor $\sqrt{c'^2}$ in turbulent flames

The value (ΔC) can be calculated if the original offgas composition is known. For pure fuels, the calculation is straightforward; for incinerators, the composition of offgas emanating from the refuse bed must be assumed. This is examined in a later section of this report. However, if ΔC can be determined, then Figure 3 can be used to determine the unmixedness factor, $\sqrt{c'^2}$. Hottel (4) has proposed that this factor be used as the characterization factor for c' to relate emissions to fluid mechanics.

To illustrate the relation of this concept to emissions, consider the time distribution of concentration, shown schematically in Figure 4, where C is the concentration of refuse bed offgases in air before combustion, C_s is the stoichiometric concentration of offgas in air, and

C_q is the concentration* of the offgas at the onset of quenching. Concentrations which exceed C_s represent regions which, if well mixed, would not have sufficient oxygen for complete combustion. Any such regions surviving until the gases leave the furnace will contribute to the total emission of combustibles. Likewise, concentrations below C_q represent regions where combustion has been quenched as a result of low temperatures. If these regions contain combustibles, then they too contribute to emissions.



*Expressed as fraction reduced to unreacted constituents.

Figure 4 Concentration fluctuations.

The source of emissions and the unmixedness factor come together when one notes that the unmixedness factor, $\sqrt{c'^2}$, is the standard deviation of concentration fluctuations about the mean concentration. If the distribution is gaussian, as Hawthorne et al (4) suggest, then $\bar{C}$ and $\sqrt{c'^2}$ are sufficient to determine the entire distribution. Furthermore, one can determine C_q and C_s from the composition of the offgas, so that the amount of gases within the concentration distribution falling outside the region from C_q to C_s can be estimated. The techniques for control then become much better defined.

From the analysis of the two emission-producing conditions discussed above, it is clear that combustible emissions can be minimized or avoided if the concentration fluctuations

*Since concentration and temperature are directly related, a quench temperature can be represented by a corresponding concentration.

are maintained entirely between C_s and C_q . This can be done either by adjusting the local air-to-fuel ratio or by increasing the level of turbulence. These two approaches, illustrated below, in Figures 5 and 6, are the province of microscale mixing.

With respect to the stoichiometry, consider the distribution of fluctuations in fuel concentrations shown in Figure 5. In distribution A, the mean concentration is close enough to C_q so that a large portion of the distribution lies within the quench zone and emissions resulting from low furnace temperatures are expected. In distribution C, on the other hand, the mean concentration is close enough to C_s so that a large portion of the distribution has a concentration greater than C_s and emissions will result because of the mixing limitation. Clearly, the emissions can be minimized, as in the case of distribution B, by adjusting the mean concentrations so that the portion of the distribution outside of the range from C_q to C_s is minimized.

The above example illustrates that the value for $\bar{C}$ can be shifted between C_s and C_q without changing the level of turbulence in the system simply by adjusting the excess air in the furnace, i.e., by controlling the furnace temperature. Of the two variables used for controlling emissions, this one is by far the most useful.

The effect of the level of turbulence is shown in Figure 6. The greater the level of turbulence, the narrower the frequency distribution becomes, increasing the probability of operating the furnace with no concentration fluctuations outside the combustion zone range. In distribution A, the mixing is poor; it is increasingly better in distributions B and C, respectively. The limiting case, of course, is that of infinite mixing (distribution D), resulting in zero emissions for all concentrations between C_q and C_s .

Fluid Mechanics and the Three T's — A considerable amount of research has been devoted to the study of turbulent mixing, primarily based on the statistical theory of turbulence as applied to the aeronautics industry. Correlations between velocity fluctuations behind a grid in a wind tunnel and stream parameters have been thoroughly studied. As a result, the decay of eddies (represented by eddy velocities) with respect to time is both empirically and theoretically described. This work is reviewed in References 5 through 8.

Relying upon the familiar similarities among heat and mass and momentum transfer, Corrsin (5) suggested the concentration analogy to velocity decay, reasoning that fluctuations in concentration and velocity both follow the same equations. The fundamental decay equation is:

$$\sqrt{c'^2} / \sqrt{c_0'^2} = e^{-\beta t} \quad (21)$$

The factor β is defined by the equation:

$$\beta = k (\epsilon/L^2)^{1/3} \quad (22)$$

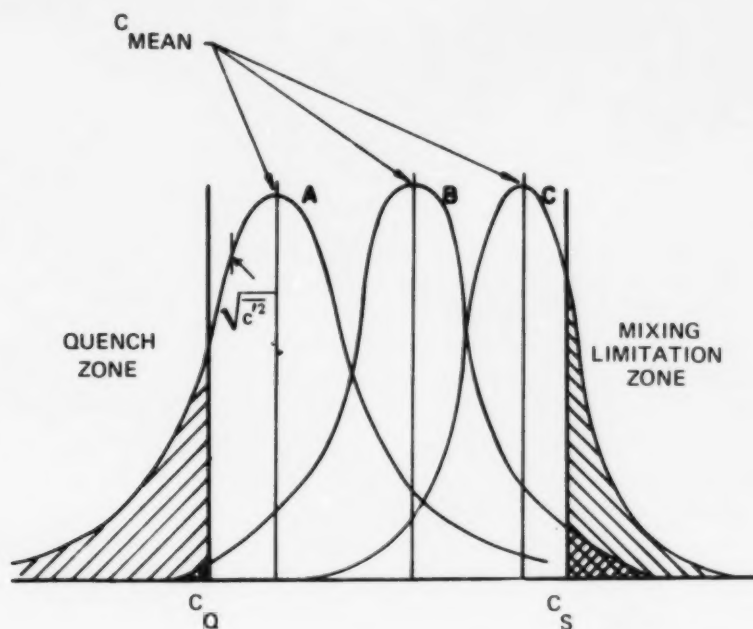


Figure 5 Effect of stoichiometry on emissions for a fixed turbulence level.

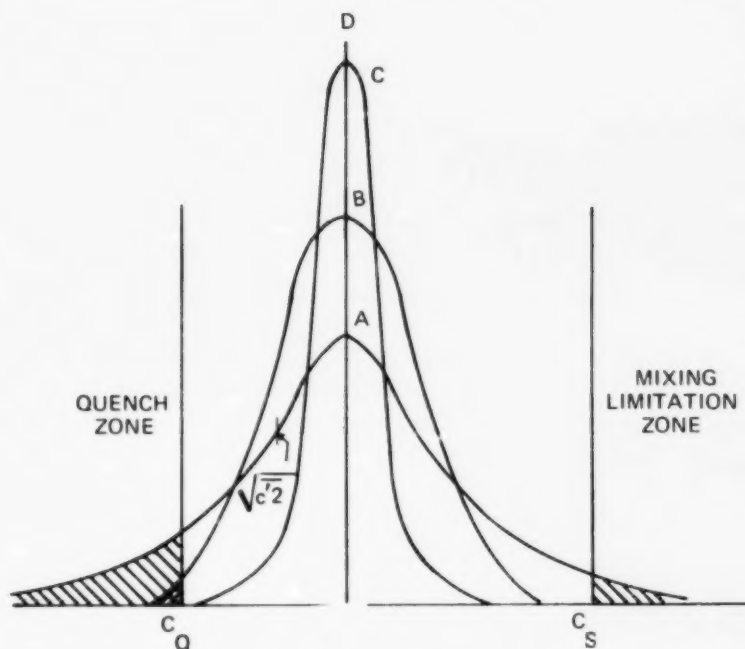


Figure 6 Effect of turbulence level on emissions for a fixed stoichiometry.

where

- L = the characteristic dimension of the system,
- ϵ = the rate of input of mixing energy per mass of system, and
- k = 0.19, as determined from turbulence experiments in wind tunnels [5].

Since the term $\sqrt{c^{7/2}}$ can be related directly to the emission factor of the furnace (through the stack gas analysis), this equation represents the relation between emissions and both time t and turbulence as characterized by β .

The effect that micromixing of this nature has on emissions is shown schematically in Figure 7. In the first diagram, the fluctuation in fuel concentration with time is shown. Because concentration fluctuations will eventually decay (according to the Corrsin model), then as long as the mean concentration $\bar{C}$ falls between C_s and C_q , the entire distribution of fluctuations will eventually fall into the combustion zone and the emissions will be eliminated. To promote this change more rapidly, one can follow two different courses of action:

- Shift $\bar{C}$ relation to the constraining boundaries C_s and C_q , or
- Increase βt to achieve a faster rate of decay in the fluctuations.

The first approach to eliminating emissions on the microscale is shown in the second diagram of Figure 7. Whenever the mean concentration is close to C_s (or alternatively to C_q), then the emissions will tend to be high. However, by adjusting the mean concentration, a greater proportion of the fluctuations will occur within the combustion zone and the emissions will be diminished. Since the concentration is directly related to the temperature, the importance of maintaining the temperature within the combustion range is clearly seen. This is the primary effect of temperature on emissions. The second approach focuses on achieving more complete decay of the concentration fluctuations.

In the third diagram, a typical concentration fluctuation for a relatively unmixed system would appear as the solid lines and would result in a given amount of emissions at a specified concentration (or temperature) and βt . By maintaining the mixture at the same concentration, but improving the level of mixing, the fluctuations can be reduced as shown by the dotted profile so that the emissions can be reduced or eliminated. The same argument also applies to increases in residence time at constant temperature and degree of turbulence. Although the singular importance of adjusting stoichiometry is crucial to the reduction of combustible emissions, these secondary relationships between temperature and both turbulence and time have also been noted in order to complete the quantitative description of the interrelationships among the three T's.

Application of Micromixing Models to Incineration – The previous two headings of this section dealt with the development of a micromixing model to correlate emissions data

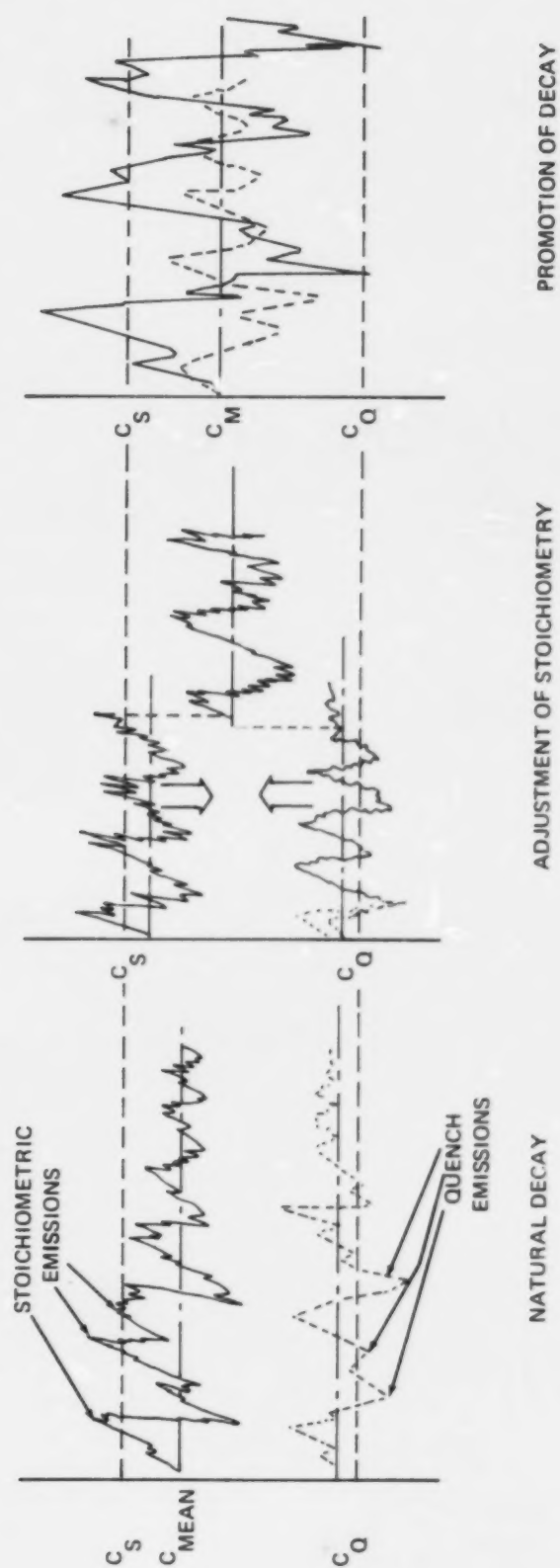


Figure 7 Overall effect of micromixing on emissions.

with furnace operating parameters, particularly the three T's. The two parts of the model that were presented are:

- Use of the unmixedness factor, $\sqrt{c^2}$ to relate emissions data to fluid mechanics properties of the system; and
- Use of Corrsin's Equation to relate the fluid mechanics properties to time, temperature, and turbulence.

In this section, we have outlined the steps required to apply both parts of the model to incinerators.

Throughout this part of the report it is necessary to understand the combustion phenomena at work within the incinerator. In the previous mixing study (2), we developed a model to characterize the bed burning process. During the course of this work, it became more and more apparent that the characterization of bed combustion was not one of the primary factors necessary to understand and control combustible emissions. However, the bed model was often necessary to estimate the value of the parameters used in the mixing models. The development of the bed burning model is given in Appendix E. Some of the results of this development are summarized below.

Within the refuse bed, temperatures are high enough to assure complete burning of pyrolysis or gasification products that come into intimate contact with oxygen. More likely, however, there will be insufficient oxygen or poor mixing so that combustible gases will be given off.

The offgas contains two distinct constituents: (1) combustibles and combustion products, and (2) air. The composition of the combustibles and combustion products is controlled by the water gas shift equilibria. The composition can be estimated (as shown in Appendix E) if the initial value of CO/CO₂ is known. Typical compositions are shown in Table 1 for different values of CO/CO₂. These are also shown graphically in Figure 8.

The actual value of CO/CO₂ was determined empirically to be 0.21 (range 0.02 to 0.6), based on test data taken directly over the bed of an existing incinerator. The data are presented in Section VII.

The offgas from the refuse bed also contains oxygen that has not yet been consumed. Since oxygen and combustibles do not co-exist in a well mixed state, the oxygen available in the offgas must be contained within a pocket of unmixed air. The fraction of such air in the offgas is simply the ratio of available oxygen in the sample (in percent) to 21%. In the tests reported in Section VII, the incinerator contained between 29% and 90% unmixed air with an average of about 50%.

Table 1. COMPOSITION OF BED OFFGAS

CO/CO ₂ →	0.25	0.50	1.0	1.5	2.0
Analysis (vol %)					
CO ₂	13.5	12.2	10.5	9.3	8.4
CO	3.4	6.1	10.5	14.0	16.8
H ₂ O	22.4	22.0	21.0	19.9	18.9
H ₂	2.8	5.5	10.5	15.0	18.9
N ₂	57.9	54.2	47.5	41.8	37.0
O ₂ (req.)	3.1	5.8	10.5	14.5	17.8
mol O ₂ /100 mol offgas					

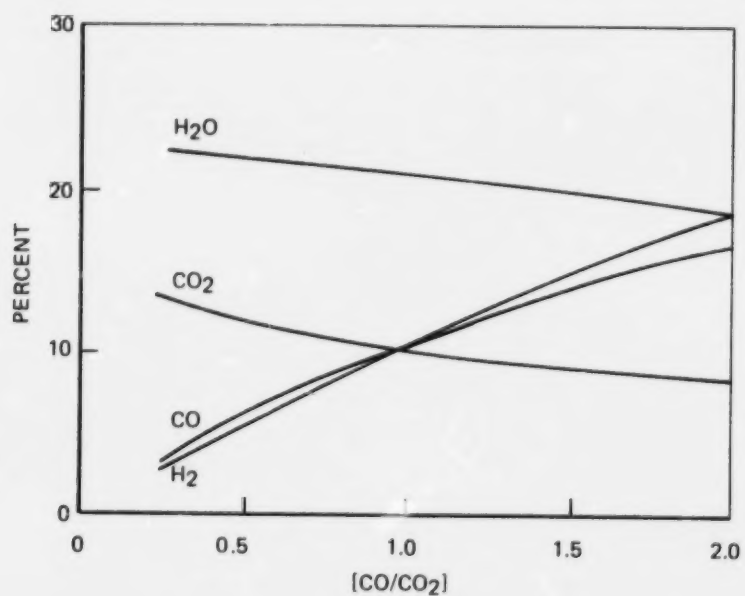


Figure 8 Offgas composition of CO/CO₂

- *Application of the Unmixedness Factor*

In Appendix B, the procedure for calculating the unmixedness factor from a stack gas analysis is given. It is also possible, given an unmixedness factor, to calculate the emission factor if the initial CO/CO₂ ratio is known. In Figures 9, 10, and 11 are shown three sets of calculations based upon different CO/CO₂ values and also different empirical relations for CO₂ vs temperature. In this work both the initial CO/CO₂ ratio and the stack gas analysis were measured. Hence it was possible to determine the value for $\sqrt{c'^2}$ two different ways and compare the results. This is also done in Section VII. The good agreement between the experimental value of $\sqrt{c'^2}$ (based on gas analysis) and the theoretical value of $\sqrt{c'^2}$ (based upon the CO/CO₂ ratio) supports the contention that emissions can be determined using the unmixedness factor approach.

- *Application of Corrsin's Equation*

To apply Eq. (21) above to incinerators, one must be able to estimate the following parameters:

- β
- t
- $\sqrt{c_o'^2}$
- Δ_c

The first two are related to the aerodynamics of the furnace mixing and also to the furnace residence time (for gases, not refuse). The latter two parameters require the knowledge of the composition of offgas emanating from the refuse bed. The estimation of each of these is discussed below.

Parameter β – For the incinerator, the only types of mixing energy that have been considered are those introduced by incoming air streams. Equation (22) can then be expanded as follows. The rate of input of mixing energy for a given air stream is calculated from the equation:

$$\frac{d(KE)}{dt} = \frac{\frac{1}{2} m u^2}{\Delta t} = \frac{1}{2} (\rho A u) u^2 = \frac{1}{2} \rho Q^3 / A^2 \quad (23)$$

If each stream is evaluated at the temperature at which it gives up its energy to create turbulence, then as a conservative estimate each jet must be evaluated at the jet temperature, while the gases coming off the refuse bed are evaluated at the furnace temperature. The energy term for a jet acting over a portion of the grate is:

$$\epsilon = \frac{\frac{1}{2} \rho_o Q_o^3 / A_o^2 \Big|_{T=T_o} + \frac{1}{2} \rho_j Q_j^3 / A_j^2 \Big|_{T=T_j}}{\rho_o V} \quad (24)$$

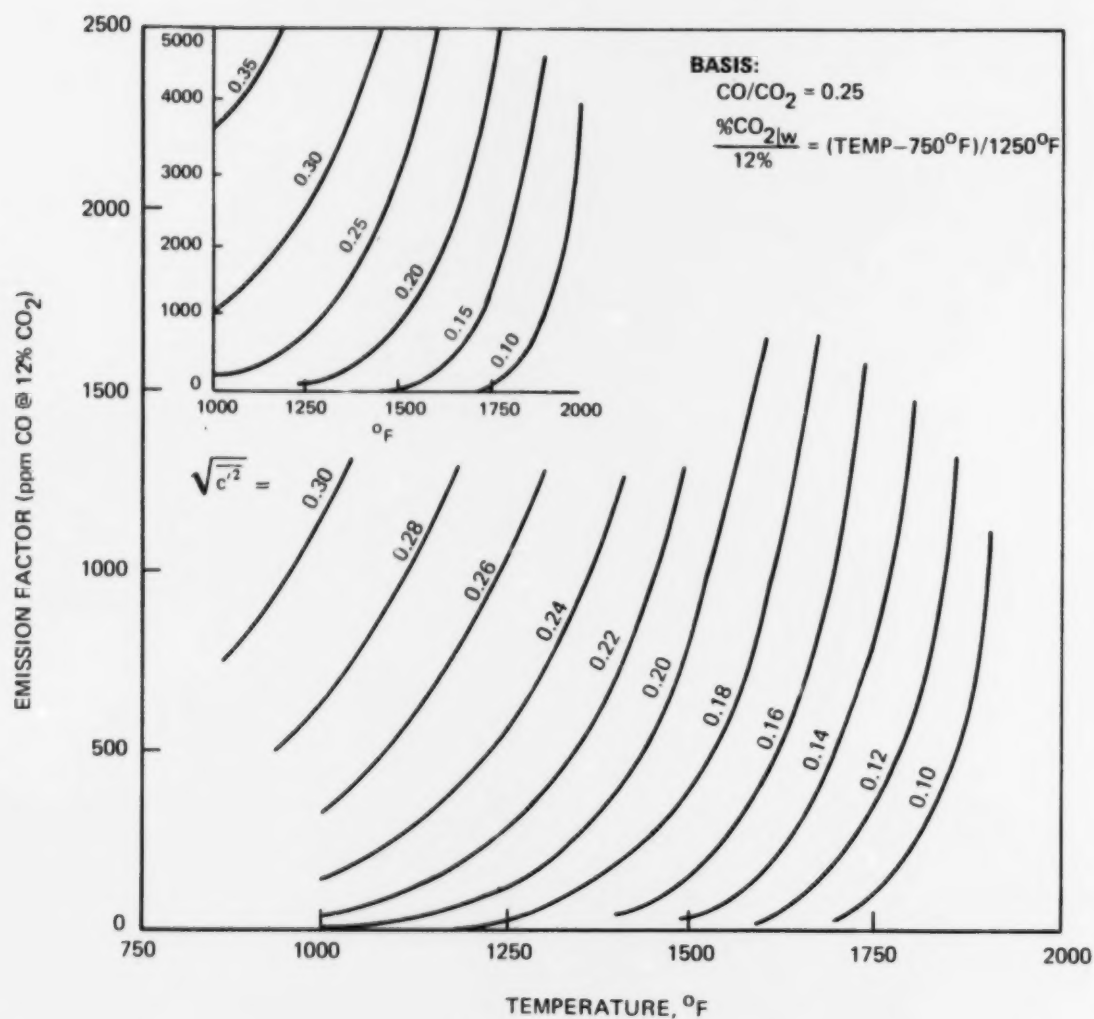


Figure 9 Dependence of emission factor on temperature and unmixedness factor.

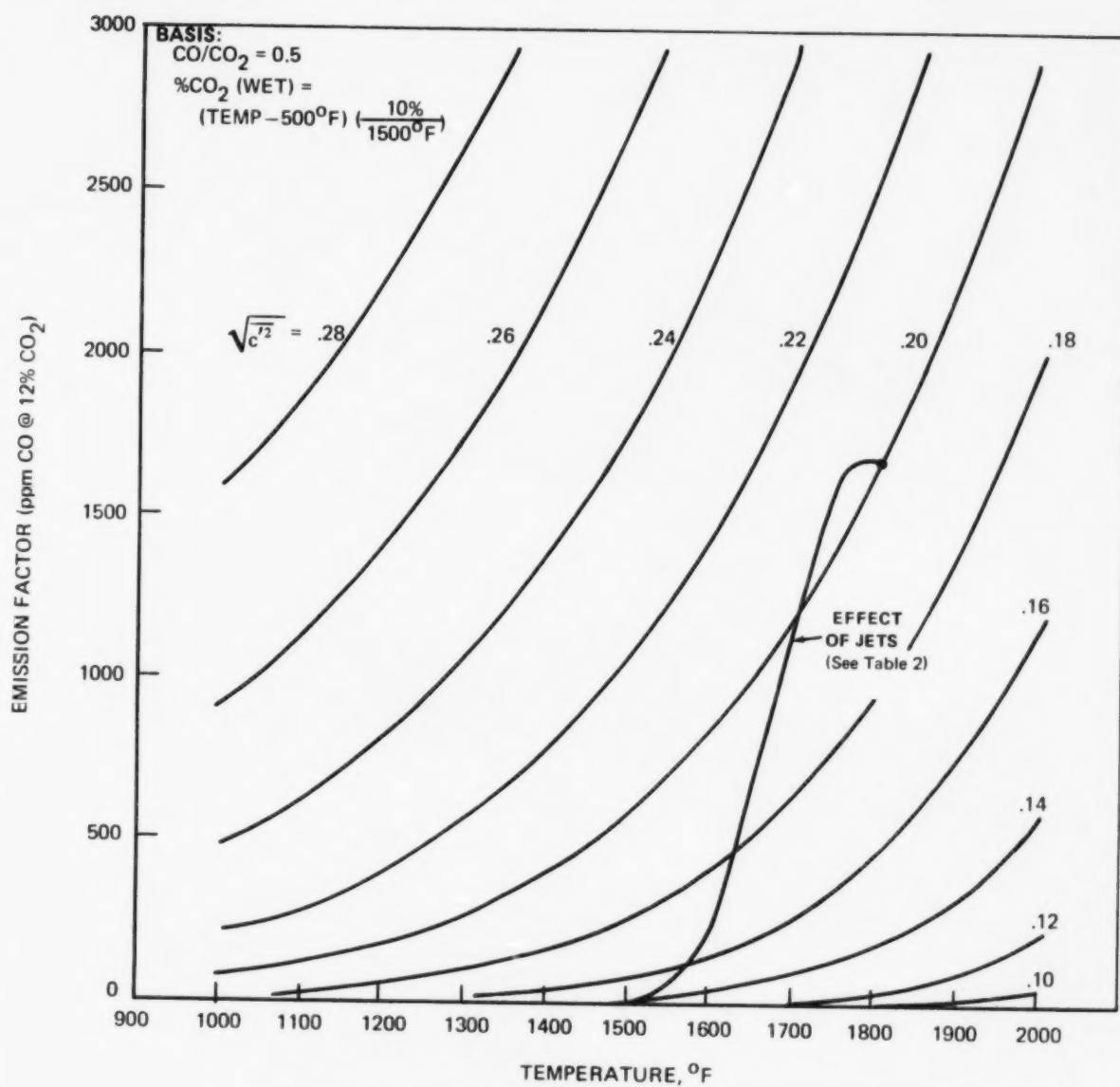


Figure 10 Dependence of emission factor on temperature and unmixedness factor.

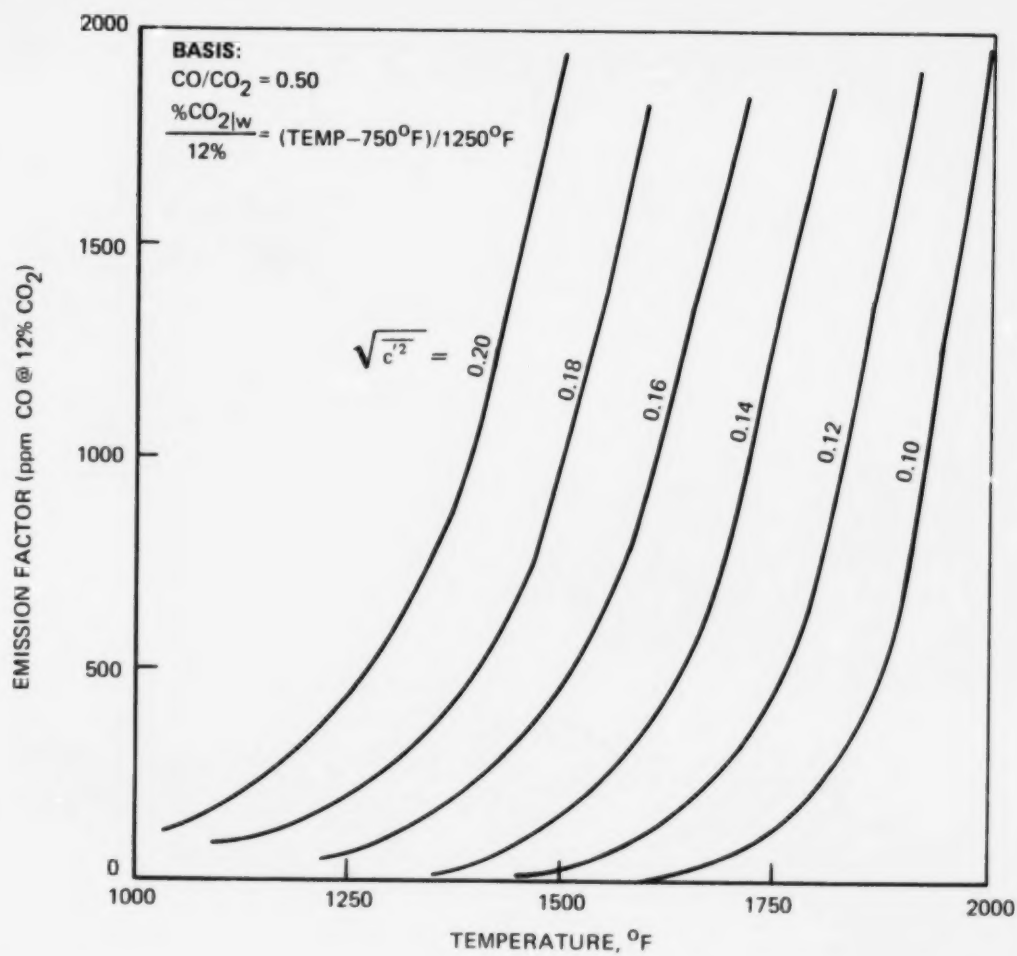


Figure 11 Dependence of emission factor on temperature and unmixedness factor.

By substituting in the ideal gas law and scaling all dimensions with the characteristic dimension L , which is taken to be the shortest dimension of the furnace (usually the furnace width), the equation for β becomes:

$$\beta = \frac{0.15 Q_o}{L^3 \eta} \left\{ \frac{1}{\gamma} \left[1 + \left(\frac{T_o}{T_j} \right) \cdot \left(\frac{A_o}{A_j} \right)^2 \cdot \left(\frac{Q_j}{Q_o} \right)^3 \right] \right\}^{1/3} \quad (25)$$

where

$$\begin{aligned} \eta &= L_f/L = \text{ratio of length to width,} \\ \gamma &= H_f/L = \text{ratio of height to width, and} \end{aligned}$$

subscripts

$$\begin{aligned} o &= \text{initial condition at top of bed} \\ j &= \text{jet.} \end{aligned}$$

If uniform mixing has not been achieved, then β will have a different value for each furnace streamline.

Parameter t - Parameter t is the residence time of combustion gases within the furnace. One method for calculating t was proposed in the section on fluid flow [see Eqs. (11) through (13)]. Another approach is to relate t to the gas flow rate Q_f and the furnace volume V_f . This could be combined with Eq. (25) above to calculate βt according to the equation:

$$\beta t = 0.15 \left\{ \gamma^2 \left[1 + \left(\frac{T_o}{T_j} \right) \cdot \left(\frac{A_o}{A_j} \right)^2 \cdot \left(\frac{Q_j}{Q_o} \right)^3 \right] \right\}^{1/3} \quad (26)$$

assuming that

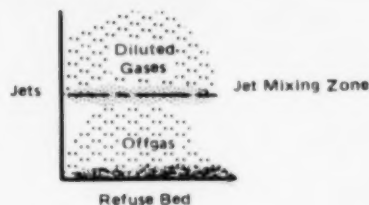
$$t = \frac{L^3 \eta \gamma}{Q_o}$$

Note that without jets $(\beta t)_o$ equals $0.15 \gamma^{2/3}$.

In this case, the dependence on furnace length is eliminated. Since macroscale uniformity is more probable across the width of the furnace or from grate to roof, but not from back to front, the use of this form of the equation is more consistent with the local character of the microscale model.

Parameters $\sqrt{c'^2}$, $\sqrt{c_o'^2}$ and ΔC - The effect of overfire air jets can be determined from a material balance of the jet mixing process using the calculated offgas composition for

refuse beds (See Appendix E). Consider for example how the composition of bed offgas is affected by dilution with high-velocity air jets. The following equations are based upon the point of jet mixing as sketched below:



Let x equal the mole fraction of offgas in the diluted gases. Then x can be estimated in one of two ways. The amount of carbon in the offgas is the same as the amount of carbon in the diluted gases. At the breach of the incinerator essentially all of the carbon is in the form of CO_2 . If a gas analysis is taken before and after the jets are turned on, and if subscript b represents the base condition (before jets) and subscript j represents the jet condition, then the carbon concentration represented by CO_2 is:

$$\% \text{CO}_2|_j = x(\% \text{CO}_2|_b) \quad (27)$$

or

$$x = \frac{\% \text{CO}_2|_j}{\% \text{CO}_2|_b} \quad (28)$$

Alternatively, a balance of total moles at the point of mixing yields the equation:

$$\frac{PQ_j}{RT_j} + \frac{PQ_c}{RT_c} = \frac{PQ_f}{RT_f} \quad (29)$$

In this equation, the variables that can be measured are Q_j , T_j , Q_f , and T_f . The defining equation for x then becomes:

$$x = 1 - \frac{Q_j/T_j}{Q_f/T_f} \quad (30)$$

The factors required to calculate $\sqrt{c'^2}$ all change with the addition of jet air. The changes in these variables are calculated using the following equation. The change of βt is best determined based upon Eq. (30):

$$(\beta t)_j = (\beta t)_b \left\{ 1 + \left[\frac{(A_f/T_f)}{(A_j/T_j)} \right]^2 (1-x)^3 \right\}^{1/3} \quad (31)$$

Addition of air will increase available oxygen according to the equation:

$$(\%O_2)_j = x (\%O_2)_b + (1-x) 21\% \quad (32)$$

Likewise the oxygen required for complete combustion after dilution is determined from the equation:

$$(O_2 \text{ Req})_{j_o} = (O_2 \text{ Req})_{b_o} \cdot x \quad (33)$$

Dividing Eq. (32) by Eq. (33) and combining terms, one obtains the following:

$$\left(\frac{\%O_2}{O_2 \text{ Req}} \right)_{j_o} = \left(\frac{\%O_2}{O_2 \text{ Req}} \right)_{b_o} \left[1 + \left(\frac{1-x}{x} \right) \frac{21\%}{(\%O_2)_b} \right] \quad (34)$$

Note that Eq. (34) applies only to the region of jet mixing (subscript o), not to the entire streamline. A similar approach is taken to determine Δy . In Appendix B it is shown that:

$$\Delta y \cong y_s \left(1 - \frac{\%CO_2|_b}{\%CO_2|_s} \right) \quad (35)$$

where y represents the mole fraction of original combustibles (before air entrainment within the bed). The addition of jet air carries the dilution of the offgases even further than occurs within the bed itself. At the exit of the bed:

$$(y-y_s)_b = y_s \left[1 - \left(\frac{\%CO_2|_b}{\%CO_2|_s} \right) \right] \quad (36)$$

After the jets:

$$(y-y_s)_j = y_s \left[1 - \left(\frac{\%CO_2|_j}{\%CO_2|_s} \right) \right] \quad (37)$$

The equation for Δy_j can be expressed in terms of x as:

$$(\Delta y_j) \cong (\Delta y_b) \left(1 + \frac{y_s}{\Delta y_b} \left[\frac{\%CO_2|_b}{\%CO_2|_s} \right] (1-x) \right) \quad (38)$$

The above equations can be combined with Corrsin's Equation and Figure 3 to provide a single model equation as follows:

From Figure 3:

$$\left(\frac{\sqrt{c'^2}}{\Delta y} \right) = f(\phi_o) \quad (39)$$

Note that $f(\phi_o)$ is given in Appendix F.

Combining this with Eqs. (34) and (38) yields:

$$\frac{(\sqrt{c^{T_2}})_j}{(\sqrt{c^{T_2}})_b} = \left[1 + \frac{y_s}{\Delta y_b} \left(\frac{\%CO_2|_b}{\%CO_2|_s} \right) (1-x) \right] \left[\frac{f(\phi_{j_o})}{f(\phi_{b_o})} \right] e^{-[(\beta t)_j - (\beta t)_b]} \quad (40)$$

The jet effect can be graphically illustrated in the following way. Figure 10 can be replotted (Figure 12) to show emission factor vs $\sqrt{c^{T_2}}$ (rather than temperature). Then Point A on Figure 12 represents some base condition as determined in the breeching of the furnace. The initial base condition can be determined by applying Corrsin's equation in the form:

$$(\sqrt{c^{T_2}})_{b_o} = (\sqrt{c^{T_2}})_b e^{-\beta t} \quad (41)$$

The initial point can be located on Figure 12 by assuming that the initial and breech conditions are at the same temperature, i.e., the decay process is isothermal as shown in Figure 12 (Point B). One could also apply the model by specifying the initial temperature rather than making the isothermal assumption. The initial jet condition can be determined by noting that the emission factor does not change with dilution, but remains constant in going from the base to the jet condition. Because of the isothermal assumption, the factor $(\sqrt{c^{T_2}})_{j_o}$ is the value determined at the emission factor of the base condition and the temperature of the jet condition, i.e., Point C. The decay of $(\sqrt{c^{T_2}})_j$ is also isothermal, so that the final value will lay somewhere along the jet isotherm, Point D, for example. The magnitude of decay can be determined from the equation:

$$\left[(\sqrt{c^{T_2}})_{j_o} - (\sqrt{c^{T_2}})_j \right] \cong \left[(\sqrt{c^{T_2}})_{b_o} - (\sqrt{c^{T_2}})_b \right] \left\{ \frac{(\sqrt{c^{T_2}})_{j_o} \cdot (1 - e^{-(\beta t)_j})}{(\sqrt{c^{T_2}})_{b_o} \cdot (1 - e^{-(\beta t)_b})} \right\} \quad (42)$$

This equation is the basis for predicting the effect of jets on combustible emissions.

Estimated Effect of Jets on Incinerator Emissions

The effect that jet mixing has on combustible emissions follows from Eq. (40) above as seen in the following example.

Consider a furnace initially at 1800°F, having an unmixedness factor of 0.20 and with βt equal to 0.25. The relationship between emissions, temperature, and $\sqrt{c^{T_2}}$ is assumed to be that of Figure 9. The empirically derived relationship between temperature and %CO₂ is:

$$\frac{\%CO_2}{10\%} = \frac{x(\%CO_2)_b}{10\%} = \frac{T - 960^\circ R}{1500^\circ R} \quad (43)$$

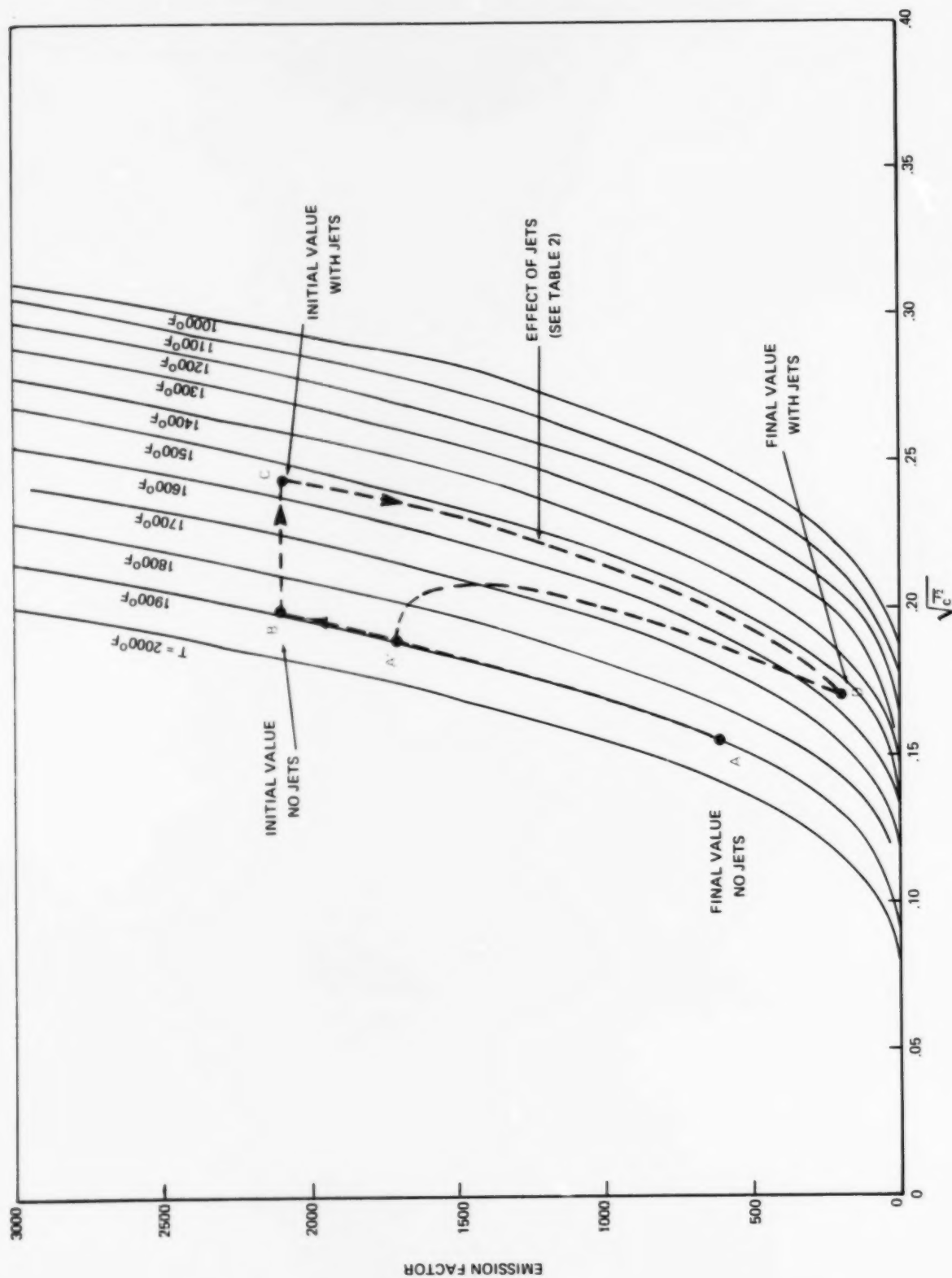


Figure 12 Dependence of emission factor on temperature and unmixedness factor (revised plot of Figure 10)

At 1800°F (2260°R), %CO₂ is 8.7%. The ratio of A_f/A_j is taken to be 150. A summary of the calculated terms for the jet conditions is given in Table 2 based upon the following calculations:

<u>Parameter</u>	<u>Source</u>
x	Given
%CO ₂	Eq. (27)
T	Eq. (43)
$\phi_{j_0} = \frac{\% O_2}{O_2 \text{ Req}}$	Eq. (34)
$\left(\frac{\sqrt{c'^2}}{\Delta y} \right)_0$	Figure 3
Δy_j	Eq. (38)
$\left(\frac{\sqrt{c'^2}}{\Delta y} \right)_{j_0}$	$\left(\frac{\sqrt{c'^2}}{\Delta y} \right)_{j_0} (\Delta y_j)$
$\frac{-(\beta t)_j - (\beta t)_b}{\sqrt{c'^2}}$	Eq. (31)
$\left(\frac{\sqrt{c'^2}}{\Delta y} \right)_0 - \left(\frac{\sqrt{c'^2}}{\Delta y} \right)_j$	Eq. (40) or Eq. (41)
	By substitution of Eq. (42)

Note that the internal consistency of the equations can be demonstrated by calculating the final two parameters listed above, using either of two equations each.

The values for $\sqrt{c'^2}$ vs temperature given in Table 2 have been plotted on Figure 12. The jets have a dramatic impact on reduction of emissions whenever x is less than 0.95. In Section VII, we have shown that other conditions could be set forth in which the jets prove to be completely ineffective or of marginal value.

Table 2. SUMMARY OF CALCULATED TERMS

x	x	T, °F	ψ_0	$\left(\frac{\sqrt{c^2}}{\Delta C}\right)_0$	ΔV_0	$\left(\sqrt{c^2}\right)_0$	$\left[1 + \left(\frac{150 \cdot T_i}{T_f}\right)^2\right]^{1/3}$	$\sqrt{c^2}$	$\Delta\sqrt{c^2}$ *
Base	1.0	1800	27	0.774	0.332	0.257	1.0	0.200	0.057
Jet	0.98	1778	28	0.765	0.341	0.261	1.004	0.203	0.058
	0.95	1740	30	0.750	0.355	0.266	1.055	0.204	0.062
	0.90	1675	34	0.725	0.377	0.273	1.356	0.195	0.078
	0.85	1609	38	0.700	0.400	0.280	1.853	0.176	0.104
	0.80	1544	43	0.680	0.422	0.287	2.442	0.156	0.131
	0.75	1479	48	0.665	0.445	0.296	3.032	0.137	0.159

$$* \Delta\sqrt{c^2} = \left[\left(\sqrt{c^2}\right)_0 - \left(\sqrt{c^2}\right) \right]$$

SECTION VI

EXPERIMENTAL PROGRAM

A large portion of this program was devoted to testing an existing municipal incinerator to:

1. verify the validity of applying the basic combustion fundamentals presented in the previous chapters to the design of municipal incinerators;
2. demonstrate the effectiveness of overfire jets in reducing combustible emissions; and
3. provide experimental data on municipal incinerator operation in hopes of promoting further technological advances in design or operation.

The test equipment procedures and data reduction methods are discussed in detail below, with the data tabulated in Appendix A.

TEST INCINERATOR

The primary piece of equipment used in the test was a municipal incinerator. Before selecting the test vehicle, we reviewed the various types of incinerators in use in the United States today. We limited our selection to grate-type incinerators, because the batch-type incinerators are being phased out around most of the country and suspension-fired units are not in sufficient use to have general applicability. We considered several local incinerators before finally selecting the incinerator of the City of Newton, Massachusetts.

The Newton incinerator is a 500-TPD capacity plant consisting of two identical 250-TPD traveling grate furnaces as shown in Figure 13. The incinerator configuration is typical of many found throughout the country today. The refuse arrives at the facility and is weighed before being dumped into the loading pit. A fine water mist is periodically sprayed over the refuse to control the dust. Unfortunately, much of this water settles to the bottom of the pit so that the refuse there is very moist and burns poorly. Both furnaces are fed by a 3.5-ton traveling bridge crane provided with a 2-yard clamshell bucket. The charge chute to each furnace is water-cooled and gravity-fed. During operation, the crane operator is reasonably careful in blending the various types of refuse to obtain a uniform mixture. Even so, large fluctuations in the refuse composition are seen so that consistent operation on a particular type of refuse is difficult to maintain for more than one or two hours. In collecting data on emissions, we preferred test runs lasting for as long as 4 hours, but in many cases changes in refuse composition or other adverse conditions forced a change in the operating conditions of the incinerator.

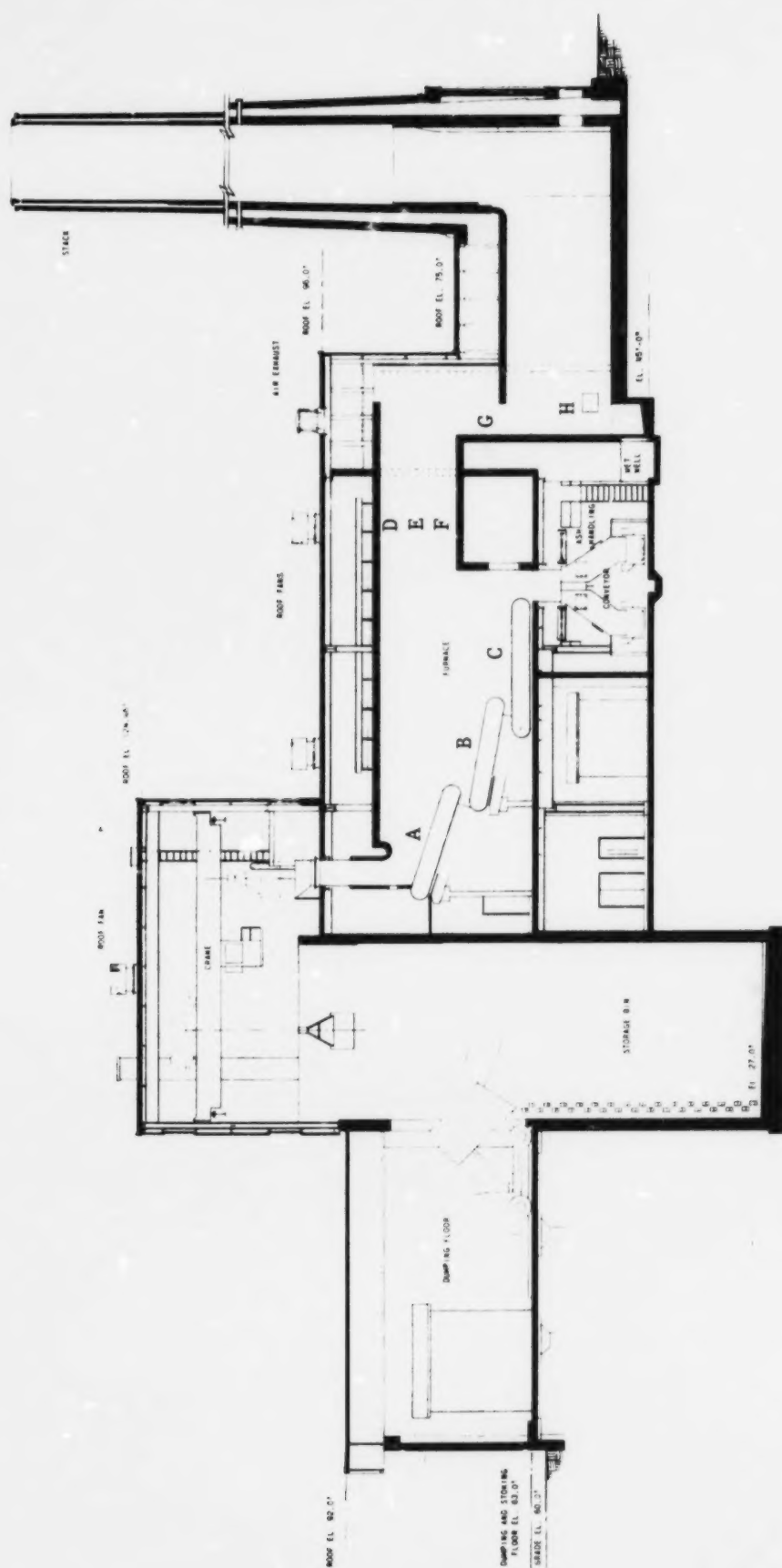


Figure 13 Cross section of Newton 500-TPD Incinerator showing sampling locations.

Furnace

Each of the rectangular furnaces is lined with conventional refractory, except for the lower portion of the wall about three feet above each grate which is lined with a silicon carbide brick to resist abrasion and slagging. The distance from the grate to the roof at the charging end is approximately 10 feet; at the drop-off end it is approximately 30 feet.

The furnace has three traveling grates to convey the refuse through the incinerator. The initial grate is approximately 11 feet long and is inclined to facilitate radiative transfer of heat to dry the incoming refuse. Hence, it is designated the drying grate. The refuse is drawn from the feed chute into the furnace through an opening 4.2 feet high by 8 feet wide. The feed rate of refuse into the furnace can, therefore, be determined by the speed of this first grate and the density of the refuse. For an average density of about 15 lb/ft³, a grate speed of 40 ft/hr results in a feed rate of 250 tons/24 hour day, which is the design capacity. The grate itself consists of rows of overlapping keys through which air is passed both to promote burning and to cool the burning surface. The keys are made of cast iron to resist high temperatures. The rack bars holding the keys are made of mild steel. Throughout the program conditions often existed that, if not properly adjusted, resulted in overheating of the steel portions of the grate. Since a severe overheating could have caused damage to the grates, these conditions were abandoned whenever the grate temperature began to rise abnormally.

At the end of the drying grate, the refuse tumbles on to the second or burning grate. The vertical distance between the two grates is approximately 5 feet which is sufficient to promote a certain amount of refuse mixing, which tends to expose additional unburned refuse to the flame. The second grate is 16 feet long and slightly inclined. Its speed can also be controlled as can the speed of the third grate. In general, the second grate is operated at a somewhat faster speed than the drying grate because past experience has shown that when the grate speed is equal or slower than the drying grate, the latter tends to overheat. The most probable explanation for this phenomenon is that the slower grate speed of the lower grate results in a buildup of refuse directly under the nose of the upper (drying) grate. Then, either because of a restriction in the air leaking into the furnace which normally cools the nose of the grate, or because of increased heat generation at that point on the grate, the grate overheats. Since this portion of the grate is not cooled by underfire air, the grate will return to the charge chute hotter than normal. When this occurs, the refuse on the grate eventually reaches a point at which it could be prematurely ignited. With the flame right next to the grate surface, an abnormally high heating effect occurs which is not relieved by the underfire air. This phenomenon happened several times during the course of the tests and prohibited operation for long periods of time at relatively slow burning grate speeds.

The refuse is tumbled for a second time from the burning grate to the third or burnout grate. The flame on this grate is characteristically lower in intensity than on the second grate and, in many cases, the flame has either been extinguished or is merely smoldering by the time the refuse has traveled the 16-foot length. One characteristic of this phase of the operation is that many of the slower burning items, such as heavy books or logs, are still burning when they reach the end of the grate. To provide a longer residence time for the furnace, this grate was typically operated at speeds equal to or slower than the second grate. Unfortunately, at slower grate speeds the residue tended to build up, restricting the air flow through the bed and, consequently, reducing the combustion intensity. A consistent problem throughout the test was the development of methods for operating this grate to achieve good burnouts, while at the same time maintaining temperatures above it that would be well above the quench temperature. The latter is extremely difficult and consequently temperatures over this grate as low as 800°F were not uncommon.

When the refuse reaches the end of the burnout grate, its ashes fall into a water-filled hopper where they are quenched. The residue is next transferred to a truck by means of a continuous-chain flight conveyor, and eventually is used as landfill. The overall volume reduction is typically between 80% and 90%.

Combustion Air System

The primary source of combustion air is a forced draft fan rated at 40,000 cfm (6 inches SP) and fitted with an inlet vane control. The underfire air is supplied to the grates by three separate ducts running from the main fan discharge to the air plenum located between the upper (forward) and lower (return) sections of each of the grates. Each plenum is divided into two sections, with the distribution controlled by a manual damper at the entrance to each system. Air is forced through the upper portion of the grate, passing through the keys and cooling the burning surface. The boxes are not particularly well sealed, so that high static pressures in the wind boxes cannot be attained. For this reason the amount of air that can be forced through the refuse bed is severely constrained, particularly in the drying grate. The third grate has very low refuse bed depths at normal or high speeds and can accept relatively large portions of air without difficulty. This configuration often results in high excess air over this grate and consequently low temperatures.

The overfire air is taken from the fan discharge using only one large duct with a single manual damper. The overfire air is introduced into the furnace through three banks of roof jets consisting of five 6-inch diameter jets each. The primary purpose of this overfire air system is to cool the overfire region. The overfire mixing system, installed for emission reductions, had a greater number of jets operating at higher jet velocities.

Additional air is used to cool the silicon carbide portion of the wall to prevent slagging. The air enters the furnace at approximately 3 feet above the grate level. The forced draft fan supplying this air is rated at 4500 cfm (6 inches SP). This source amounts to less than 10% of the total air flux into the furnace.

By far the largest source of air is in the form of uncontrolled leakage into the furnace. The leakage rate was calculated by material balance and, typically, ran as high as 30,000 cfm. A large fraction of this air is drawn in from beneath the grates and is a natural source of cooling for both the grate and the overfire region.

Stack Gas Removal

Each furnace is equipped with a fly-ash removal chamber containing water sprays and fire brick baffles. The hot combustion gases pass through the breeching area of the incinerator into these chambers. The gases are cooled to about 350°F and the large fly ash is removed. Water from the baffles is collected in a sump, clarified, and recirculated. The baffle chambers merge directly into a single natural draft chimney which is 175 feet high and designed for the gas flow required to accommodate the ultimate capacity of both furnaces. Because the cooling of the baffle system maintains the exit temperatures in a fairly narrow temperature range of approximately 250° to 450°, the chimney draft that results is relatively constant. For this reason, both the draft pressure in the furnace and the gas flow rate through the furnace remain constant. Hence, the air distribution could be somewhat controlled, but not the flow rate. This constraint was a constant source of operational difficulty throughout the tests.

Incinerator Control

To control the incinerator at the proper conditions, additional instrumentation was required to augment its present monitoring system. The grate speeds were determined with the existing rate meters, and the speeds were verified to ensure that the meters were reading properly. Pitot tubes were installed in each of the underfire air ducts and in the main overfire duct to record the air input into the furnace. The total air volume was checked with a pitot tube reading at the inlet duct to the main fan. In addition, the leakage air was estimated using a hot-wire anemometer to determine the air velocities at the most probable points of entry.

INCINERATOR MODIFICATIONS

To follow the strategy for eliminating combustible emissions on both the microscale and the macroscale, one of the furnaces was modified with an overfire jet system. Based on information gathered during the initial test of this program, it was clear that the gases from grates 1 and 2 were very hot and oxygen-deficient, while the gases coming from grate 3 were at or below the quench temperature. To demonstrate the design principles and effectiveness for jet systems, we installed sidewall jets over grate 2 to satisfy the stoichiometry sufficiently and also to achieve sufficient penetration. The forced-draft fan that was selected had a design capability of 15,000 cfm at 6 inches SP. To achieve the greatest flexibility and control, the fan was equipped with both inlet vane control and an outlet damper.

The duct network consisted of 16-gage welded tubing and channeled air from the outside through the fan to a Y-connection, where the air was split into two streams; each stream was delivered to a jet manifold located on opposite sides of the furnace. The air distribution was monitored using a pitot tube in the main inlet duct and pitot tubes in each of the two ducts feeding the jet manifolds. In addition to the controls on the fan, a butterfly damper was installed in each of the two manifold ducts.

The manifold system consisted of a long header off which 10 duct sleeves ran, carrying air to the 10 jets on each side of the furnace. Each tubular sleeve had a 5-inch diameter and was equipped with both a 1/8-inch velocity pitot tube and a butterfly damper for velocity control and monitoring. Because of the experimental nature of the program, the duct sleeves were connected to the jets by flex-hose connections. During the course of the tests, the flex tubing caused many maintenance problems. Thus, this type of construction is not recommended for permanent installations.

The jet nozzles were constructed of 5-inch, schedule 40, mild steel pipe, inserted through the wall of the furnace to within 2 inches of the inner refractory wall. The external portion of the pipe was connected to the duct sleeves by means of a flanged connection. In installing the jets, the location for each jet was marked on the inside of the furnace, and the hole was made using standard techniques. The jet pipe was inserted through the wall, welded to the outer furnace shell, and surrounded with a high-temperature insulation. The inner refractory wall was then molded around the jet using a plastic refractory. During the course of the tests, slag buildup around the jets was quite common. Therefore, a permanent jet installation should use precut silicon carbide refractory or other slag-resistant material to avoid this problem.

CONDUCT OF TESTS

The composition of refuse that is burned in the average municipal incinerator varies from hour to hour and day to day, making it difficult to arrive at any consistently reproducible test conditions. During the operation of the Newton incinerator, these variations could be seen in breeching temperatures fluctuating as much as 500°F within a period of 30 minutes; variation in the breeching temperature of $\pm 100^\circ\text{F}$ was considered a relatively stable condition. It was therefore difficult to measure precisely all of the variables affecting the system.

Test Parameters

Test variables included the refuse feed rate, moisture content, the underfire and overfire flow rate and distribution, and the refuse residence time. Unfortunately, the test variables are interrelated and are often dictated by the permissible or possible conditions under which the incinerator can be operated. Each of these parameters is discussed in detail below.

Feed Rate

The refuse feed rate is controlled by the grate speed of the drying grate and it typically runs between 35 ft/hr and 55 ft/hr, being set according to whether the refuse is relatively wet or dry. The volumetric rate in cubic ft/hr is a product of the grate speed in ft/hr and the area of the charging port (33.6 ft²).

The weight of refuse depends upon the refuse density which often varies. The density of Newton's refuse was reported previously by Kaiser's[8] to be 13 lb/ft³, but similar data, both at Newton and for other incinerators, indicate densities as high as 17 lb/ft³. In the calculation of feed rates there is an additional uncertainty due to the packing of the feed hopper; if it is packed loosely with large voids, the apparent bulk density will be lower than 13 lb/ft³. On the other hand, a tight packing will result in a higher apparent bulk density. The density also varies with moisture content, tending toward the higher values for wet refuse.

In the conditions which were used in the current work, the following feed rates were utilized:

Grate speed, [ft/hr]	35	40	45	50	55
Volume feed rate, [ft ³ /hr]	1176	1344	1512	1680	1848
Mass feed rate,* [tons/hr]	7.7	8.7	9.8	10.8	11.9
Heat release,** [MM Btu's/hr]	68-78	78-87	88-98	98-107	108-118

* $\rho = 13 \text{ lb/ft}^3$

**heating value of refuse = 4500-5000 Btu/lb, 100% burnout.

Moisture Content

Refuse will always contain moisture but it can at times have considerably more moisture than usual. Typical values range between 20% and 50%. At Newton, the higher moisture conditions result usually after a rain, or when the refuse is taken from the bottom of the pit. To control the dust, the pit is periodically sprayed with water while trucks unload. Excessive moisture drains to the bottom of the pit. At times water was observed dripping from the clamshell bucket as refuse from the bottom was charged. The effect of moisture was clearly shown during the runs with the amount of refuse moisture changing over the course of a day's run as the crane worked from the top to the bottom of the pit.

In describing refuse, we have qualitatively used the labels of "wet" or "dry," but we recognize that the absolute moisture content is not such a discrete function. We also recognize that our visual observations could be quite misleading, and that the data on the combustion products could contradict our initial moisture labels. Of the principal variables, this one is by far the most uncertain and least controllable of the operating variables.

The moisture content was allowed to vary normally, resulting in conditions for the first tests fairly evenly divided between "dry" and "wet" refuse. Some of the runs reflect a change from dry to wet during the progress of a test from beginning to end. The final series of tests demonstrating the use of overfire jets had primarily dry refuse.

Unfortunately, the refuse moisture content, which drastically affects the refuse heat value, also affects very much the allowable feed rate. Wet refuse is normally burned at a slow refuse feed rate to achieve an acceptable burnout. If higher feed rates are attempted, the ash level on the burnout grate tends to build up, and the ash conveyor system becomes overburdened. Continuous operation in this mode during some of the tests resulted in temporary shutdowns because of the strain on the ash conveyor system. On the other hand, by reducing the second and third grate speeds a good burnout could be attained, but then the furnace would overheat. Hence, wet refuse was almost always run at low grate speeds. Dry refuse, on the other hand, was normally run at high grate speeds, and we found that at low grate speeds most of the combustion was completed on the first and second grates, with little combustion taking place on the burnout grate. However, it was still possible to vary feed rates for dry refuse between 35 ft/hr and 55 ft/hr with only minor interdependencies.

Air Flow Rates and Distributions

Because of the natural tendency toward a constant furnace air flow induced by the natural-draft stack, the effects of varying air flows and distributions were difficult to detect. However, various air distributions are used primarily in the underfire air system where the air flow rate is related to burning rates. Since most of the leakage is into the overfire region, these effects were relatively unaffected.

The underfire air can be divided into six discrete streams, since each underfire air plenum has two sections. In practice, controlling dampers become clogged with ash reducing the effectiveness of control. Care was taken during tests to ensure that these dampers did operate effectively, but some reduction of control was still observed.

The quantity of air flowing under each grate was measured by means of a pitot tube. The flow was initially calibrated by complete traverses of each duct, and periodic checks were made to verify the continued accuracy of the sample point readings. While it was hoped that the underfire air would be in excess of stoichiometric air, this could not be readily achieved, particularly under the first two grates.

The distribution of air was not as controllable as desired either. Only very small amounts of air (5% stoichiometric) could be forced under the drying grate without causing the draft to drop to zero (or slightly positive pressure), forcing smoke and fire out of the furnace. The second grate underfire air generally accounted for 15 to 30% of the stoichiometric requirement, while the remaining 35-40% of stoichiometric air fed under the grate came through the third or burnout grate. Because of a combination of buoyancy effects (larger drafts at lower elevations and a lower pressure drop through the residue of the burnout grate), it was difficult to reduce the air under that grate significantly. In addition, a large amount of leakage was present around this end of the furnace.

The two existing sources of controlled overfire air are the roof jets, used primarily for furnace cooling, and the wall-cooling air. During most tests the roof jets were off, and the rate of the wall-cooling air was consistently about 4,000 cfm.

The overfire jet system described earlier could supply up to an additional 20,000 cfm. It is interesting to note that at high jet flow rates the leakage was noticeably curtailed. The various types of flow distributions common to the jet system are discussed in detail in the section covering test results (see Section VII).

The remaining flow was uncontrolled resulting from the natural tendency of the furnace to draw air from the outside. The leakage was great typically between 20,000 and 30,000 cfm, but under the right conditions this could be reduced to a very low flow rate. In fact, very high forced air flow rates would result in almost zero leakage, but would effect a noticeable increase in the amount of smoke emissions onto the operating floor. In addition, the reduction in leakage usually resulted in an increase in the nose temperature of the grates (loss of natural cooling). These factors effectively limited the amount of underfire air that could be used.

Furnace Residence Time

Each of the three grates can be controlled independently so that the residence time for the refuse is simply the sum of the residence times on each of the three grates. Under most runs the residence time was approximately one hour. Certain combinations of grate speeds, however, are not acceptable, as discussed previously, because of the tendency of the drying grate to overheat. For most runs the three grate speeds were set at the same speed, but some tests were made at relative residence times both slower and faster than the normal condition.

Test Sequence

In each test the run conditions were first established and the incinerator was allowed to reach steady-state operation before test data were taken. During the first 30 minutes to one hour, the various parameters were monitored and final checks were made on the instrumentation to ensure that no major difficulties would be encountered during the run. This waiting

period was necessary to allow the furnace to reach steady-state operation at the test conditions, and was approximately equal to residence time of refuse in the incinerator. Although two or three residence times might be desirable from a theoretical standpoint, the large variation in the refuse composition over the course of a run was enough to obscure minor deviations from the true steady state and additional waiting time was not warranted. In the initial sets of tests used for verifying the analytical methods, data were taken over a period of 4 to 5 hours. Because of improvements in the instrumentation and analytical procedures, most of the data for the jet tests were taken over a period of one-half hour on most runs. In this time two complete sweeps of the three breeching zones could be achieved, corresponding to two sequential 15-minute average moisture determinations. The gas analysis was recorded before and after the test and trends were compared to assure that the gas averages were reasonably representative of the overall operation. The results of these tests are presented in Section VII.

Input Data

During the course of each test the following data were recorded:

- The furnace control panel readings, which include the furnace breech temperature (high temperature), the stack temperature, the furnace draft pressure, the chimney draft pressure, the forced-draft fan pressure, the under-grate pressure, and the grate speed for each of the three grates;
- The air distribution and flow rate, including velocity measurements for each of the three underfire ducts and the overfire duct;
- The velocity of each jet, as well as the total flow to the fan and to each jet manifold;
- The static pressure underneath each grate (during selected runs), and
- Overall furnace conditions and specific events (taken chronologically) which might affect the test data or furnace performance, including any shutdowns, changes in refuse feedstock or furnace operating conditions, the degree of burnout, and the approximate condition of the refuse feed.

The emissions data and the data reduction techniques used are discussed below.

Emissions Data

Measurements of emission composition, temperature, and flow were made in the breech and over the bed. Further measurements of static pressures were made at selected points to allow estimation of air leakage into the incinerator. Particulate emissions were made in the stack during the initial baseline series of tests and are reported in detail in Section VII.

Output flows were calculated from the breeching area velocity data collected using S-type pitot tubes on the gas-sampling probes and the temperature measurements. Composition data were obtained to measure the overall combustion characteristics and the combustible emissions. The species selected for measurements were:

CO ₂ - carbon dioxide	H ₂ - hydrogen
O ₂ - oxygen	CO - carbon monoxide
N ₂ - nitrogen	HC - total volatile and gaseous hydrocarbons
H ₂ O - water	

In addition, during the particulate measurement samples were collected for

Cl - chloride (as hydrogen chloride),
SO _x - total sulfur oxides, and
NO _x - total nitrogen oxides.

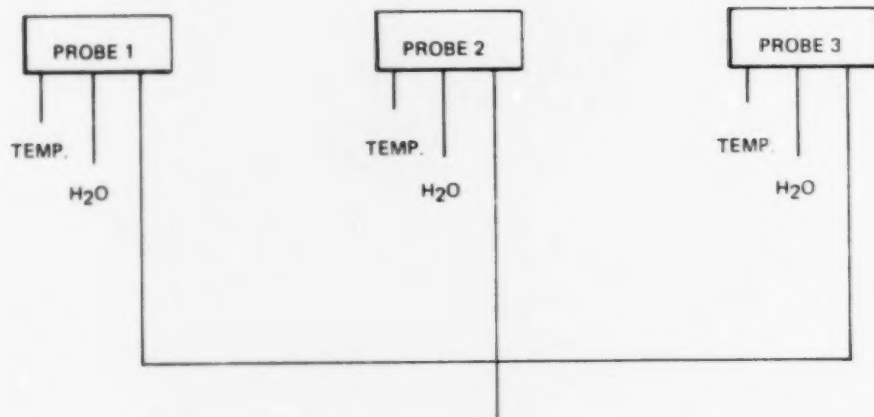
The collected particulate was also analyzed for trace metals.

Measurements were made at several points in the incinerator system. A cross section of the incinerator is shown in Figure 13. The majority of the measurements were made in the breeching at points D, E, and F. During studies of the bed-burning model, composition measurements were obtained along the bed at points A, B, and C. The continuous-stack CO monitor drew samples from point H. Initially velocity measurements were made only at the points of gas composition measurement, but in the final jet-mixing phase of the program, velocity and pressure measurements were made at points A-H as discussed later.

Initial experimental design called for operating the incinerator for an entire day (5-6 hours) at a single condition to normalize refuse-variation effects. On that basis the analysis system was designed on a time-sharing basis so that a single set of analytical instruments could be used to measure the composition of three points in the breeching or furnace. Velocity and pressure measurements were also time-shared on a single instrument, while temperatures were all recorded simultaneously on a multipoint recorder. Problems with obtaining accurate moisture analyses necessitated the use of an independent sampling line piggybacked on the gas probe for H₂O determinations. The general overall measurement approach is represented schematically in Figure 14.

Analytical Procedures

The basic sampling systems are shown schematically in Figure 15. There were three such identical systems connected to the gas analysis instruments on a time-shared basis. The probes, cyclones, and H₂O silica traps were mounted on the staging at the breeching (or by the furnace). A shed approximately 3 x 3 x 4 meters was constructed on the building floor beneath the breeching, and the gas, pressure, and temperature lines were brought to the equipment in the shed. There was approximately 10 m of sample line between the probes and the shed.



ANALYTICAL INSTRUMENTATION GASES, PRESSURE (ΔP)

Figure 14 Overall emissions measurements approach.

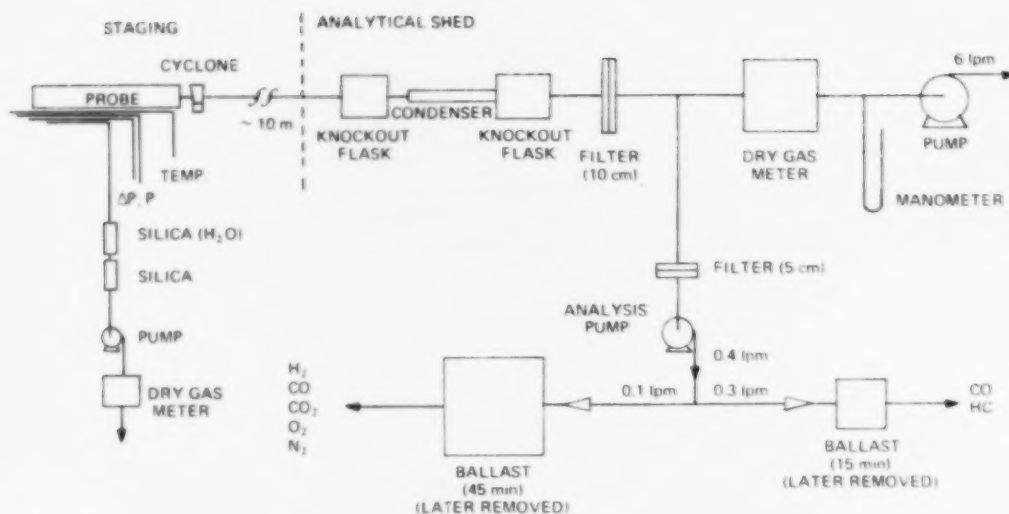


Figure 15 Sampling system.

The original intent was to make possible the collection of the H_2O for analysis by condensation and volume measurement at the end of a run. Similarly, particulate would be collected in the cyclone, lines, knockout flasks, and first filter. Our initial runs showed that the moisture content was quite low, and that only about one-half of the H_2O could be collected by condensation. Consequently, a separate sample line — not cooled — was strapped to the water-cooled gas sample probe and H_2O determined by weight pickup on 1 cm x 10 cm silica gel traps connected to a conventional gas-sampling system.

The three probes sampled gas constantly at a high enough flow rate so that we could always be assured of a representative incinerator sample, even though there was a relatively large volume in the system. The combination of a lower than desired flow rate required to prevent too high a pressure buildup on the 10-cm filter and the probe wall-effect problems eliminated the possibility of obtaining representative particulate samples for %C analyses.

Smaller pumps constantly pulled an aliquot of the gas from each of the three filtered combustion gas streams for the instrumental measurements. In the baseline series of tests, gas chromatographs (GC) were used for the H_2 , CO_2 - O_2 - N_2 analyses. One of the GC units required 15 minutes for a full cycle; therefore the portion of the gas stream (0.1 ℓ /pm) used for these analyses was passed through a 4- ℓ container having a 45-minute time constant. In this manner the analyzers could switch between the gas stream every 15 minutes and obtain a sample for analysis representative of the entire time period preceding the last analysis of that line. The more rapid CO and HC analyzers were cycled every 15 minutes, 5 minutes per sample line.

The results from the baseline studies showed that incinerator conditions really could not be held constant for 5 to 6 hours, and that observations would have to be made over a shorter time (1 hour). Consequently, the ballast cans in the gas sample lines were removed and continuous CO_2 and O_2 analyzers were substituted for the GC analyzer. Each sample line was then analyzed for 5 minutes, giving a 15-minute cycle time for all three probes for each analysis.

The gas analysis system is shown schematically in Figure 16 as used in the preferred configuration with continuous analyzers and without the ballast cans. The much greater time variation in composition than expected places a clear preference on continuous and rapid analysis, and we strongly recommend this approach over the slower batch-type analyses in all future studies.

Basically the instruments are simply switched between the three probe gas lines by means of the solenoids S1-S3 and S'1-S'3 and cycled between them every 15 minutes with 5 minutes on each probe line. The two-way solenoids allow continuous flow through the lines, so that a representative sample of the incinerator at that moment is always present.

The calibration gases were manually switched into the instruments for zero and span checks at the beginning and end of each period of data collection.

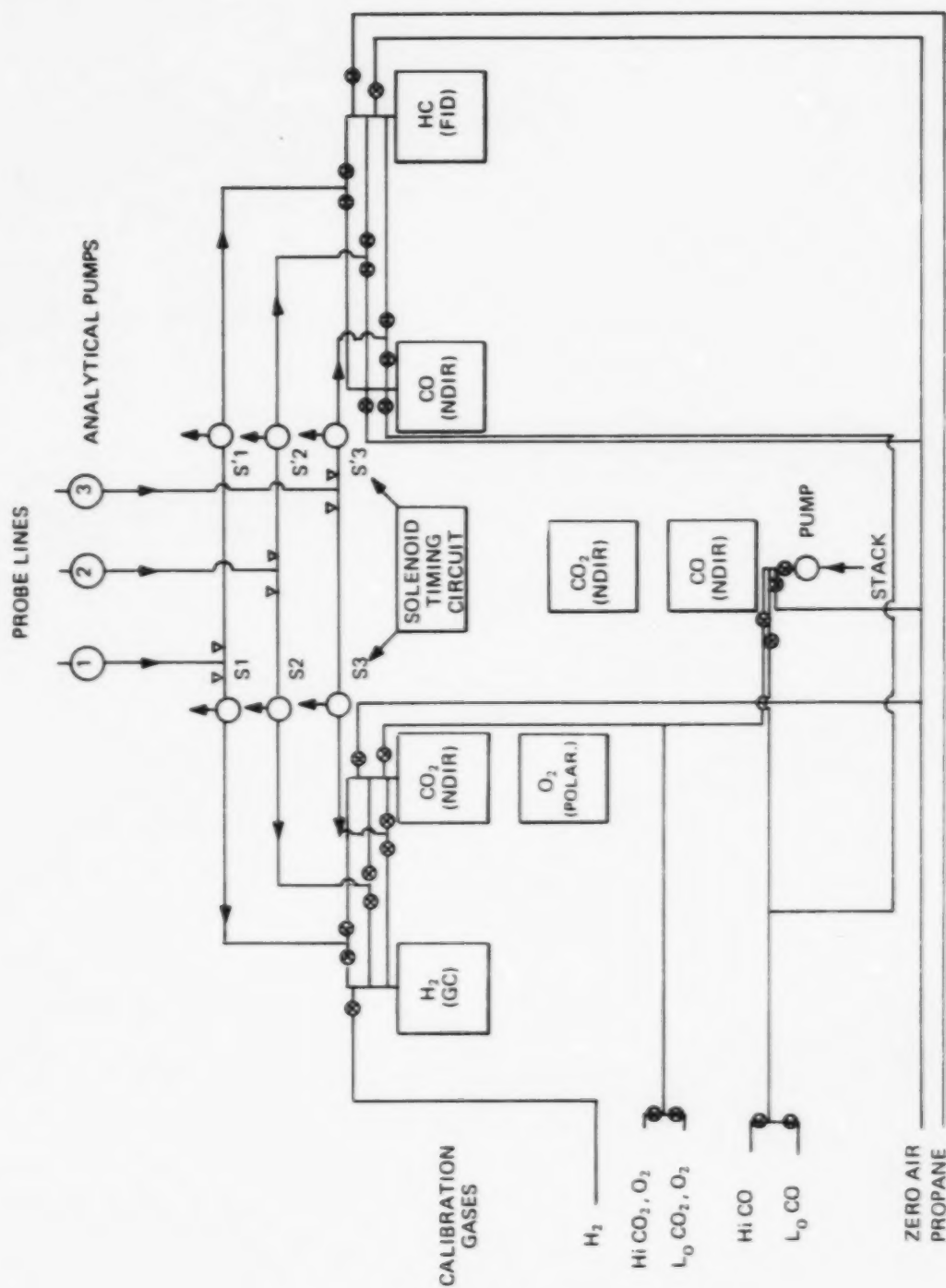


Figure 16 Gas analysis system

The starch CO/CO₂ monitor ran continuously during periods of test activity monitoring the concentration at point H in Figure 13.

The pressure measurements were made with the S-type pitots on the gas-sampling probes and various other static pressure taps in the incinerator system, as indicated in Figure 13. The pressure-measuring lines were connected through a separate timer-controlled, solenoid switching manifold system, as shown schematically in Figure 17. A set of 10 measurements was made and recorded automatically every 10 minutes. The sequence of measurements is given in Table 3. The recorded signal from the capacitance manometer was calibrated for ± 1 cm H₂O with an inclined manometer.

Table 3. VELOCITY AND STATIC PRESSURE MEASUREMENT LOCATIONS

<u>Sequence</u>	<u>Function</u>	<u>Probe</u>	<u>Probe Location[†]</u>
1	Manometer Zero Check	---	---
2	Velocity	S-Pitot	D
3	Static Pressure*	Static Tap	D
4	Velocity	S-Pitot	E
5	Velocity	S-Pitot	F
6	Static Pressure*	Static Tap	F
7	Static Pressure*	Static Tap	B or A**
8	Static Pressure*	Static Tap	C
9	Pressure Drop across Scrubber	Static Taps	G and H
10	Furnace Room - Outside Building Pressure Comparison	Static Taps	Furnace Room and Outside Building

*Reference: furnace room pressure

**Location changed from B to A for measurements on and after April 17, 1973.

† See Figure 13.

ANALYTICAL EQUIPMENT

Probe

Considerable initial evaluation and trade-off consideration went into the final design of the gas-sampling probe. The first experiments showed that an uncooled 1.2-cm SS tube could not physically withstand the 900-1000°C breeching temperatures and, more importantly, that further reaction of the gases occurred in the hot sampling tube. When evaluated

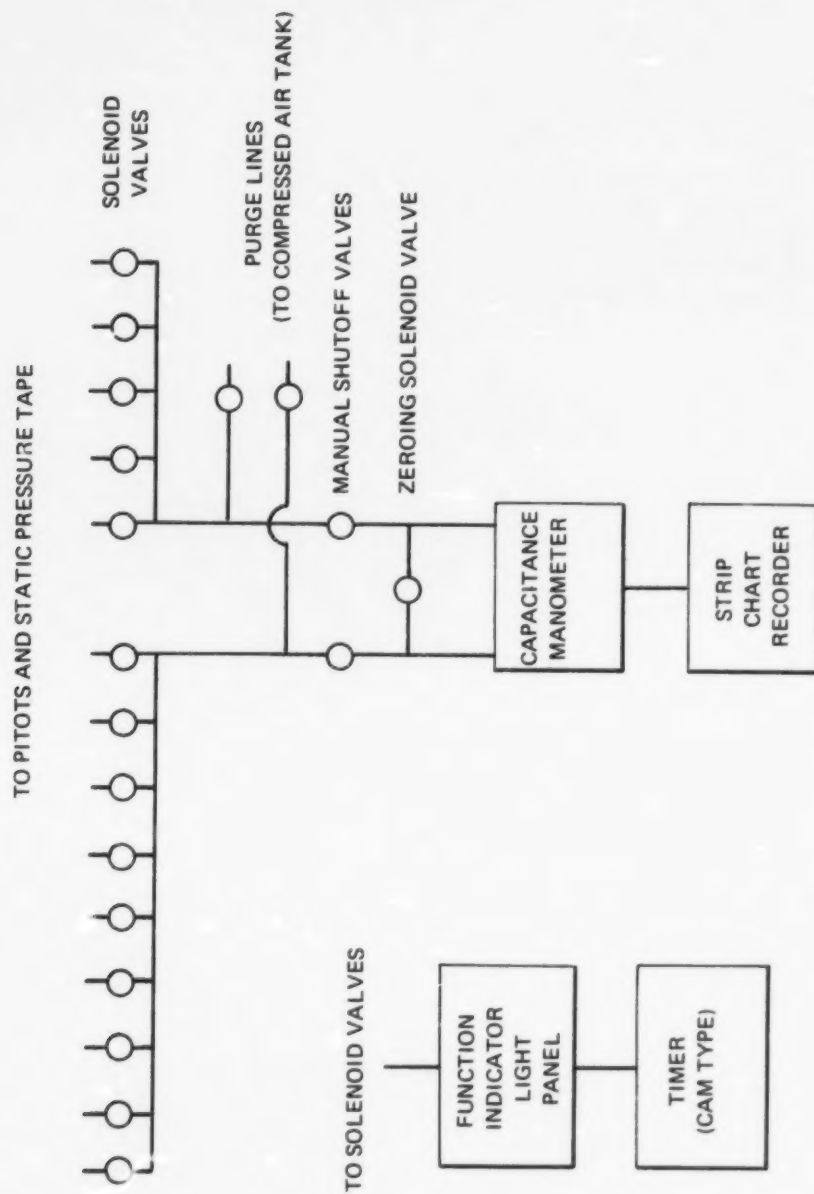


Figure 17 Static pressure and velocity (ΔP) system

materials which might physically withstand the temperatures either had an excessively high cost or a high probability of breakage. The EPA loaned us a water-cooled probe* for design evaluation, and we found that, using this probe, we could quench the combustion gas immediately after it entered the probe, and that relatively inexpensive materials could survive the incinerator environment without too much damage.

Three probes were designed and constructed as shown schematically in Figure 18 (Section AA). This type probe essentially consists of a pair of concentric tubes with an inner tube for inlet cooling water. An additional internal tube was included for transmission of the thermocouple leads and fastening the thermocouple. A glass liner was used to decrease catalyzed wall reactions of the combustion gases. The first probes had a 90-deg sample inlet so that samples, including particulate, could be obtained parallel to the flow of the gas stream. Two problems were encountered with this design. The welds in the region of the 90-deg angle could not withstand prolonged 1000°C temperatures. In fact two probes failed. Furthermore, we found that the impacting fly ash simply caked over the gas inlet and caused sampling problems. Consequently, the ends of the probe were cut off at Section AA and reconstructed as shown in Section BB. Sampling perpendicular to the gas stream eliminated the fly ash caking problem and none of the modified probes failed.

The S-type pitot tubes clamped to the probe worked well when clean, but, like the gas inlet, the upstream side of the pitot constantly became caked over and fused. Reliable pressure (ΔP) measurements could only be obtained with frequent (every hour) removal of the probe and cleaning of the pitots.

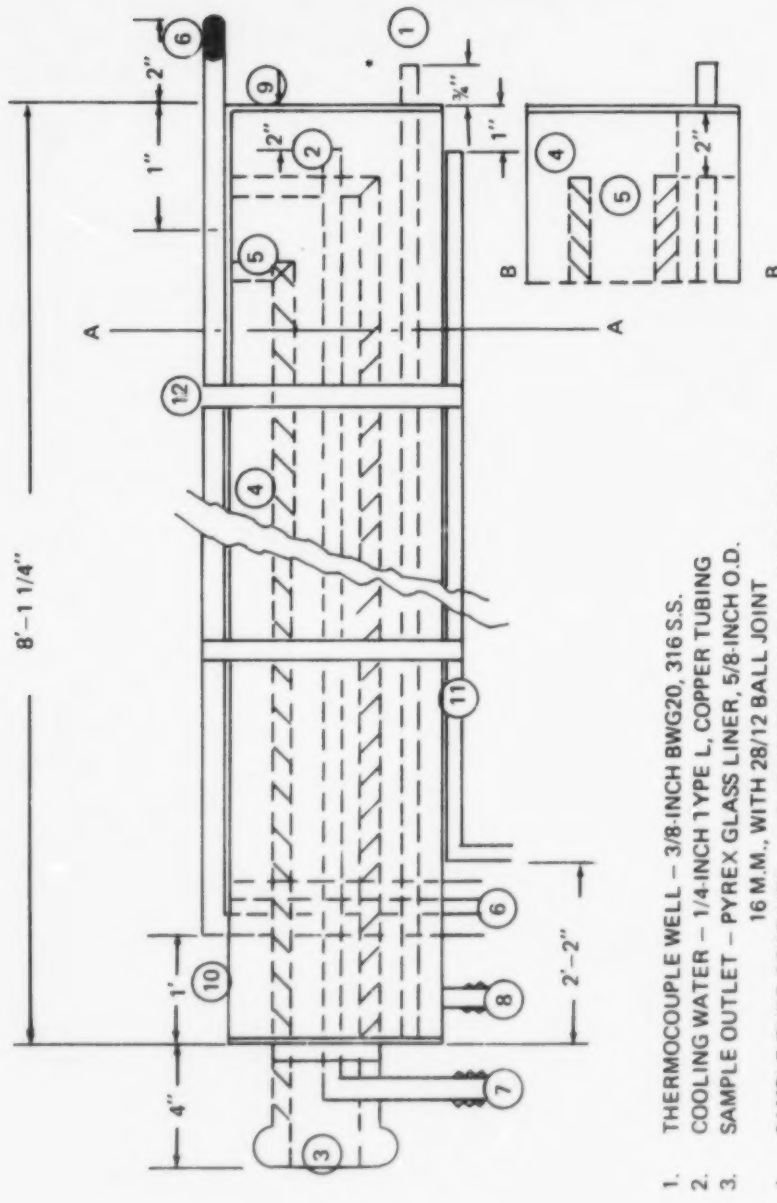
It was very important to maintain a high flow of cooling water through the probes, since even a very brief overheating caused probe sagging, linear breakage, and a run abort. In a project such as this, even the simplest things become problems. For example, we found that high-quality garden hose used to supply the probes could not be trusted and had to be carefully pressure-checked for burst strength.

Temperature measurements were made with an Omega Pt/Pt 10% Rh thermocouple housed in a ceramic casing. The thermocouple was held on the end of the probe at position 1 in Figure 18 as shown in Figure 19. The details of the radiation shield used to protect the thermocouple from the incinerator radiation and assure accurate gas temperatures are also shown.

Some measurements of "temperature only" were made over the burning refuse bed, while gas compositions were determined in the breeching. For these measurements, three simpler water-cooled probes were made similar to the gas probes, but they contained only the line for the thermocouple and the radiation shield.

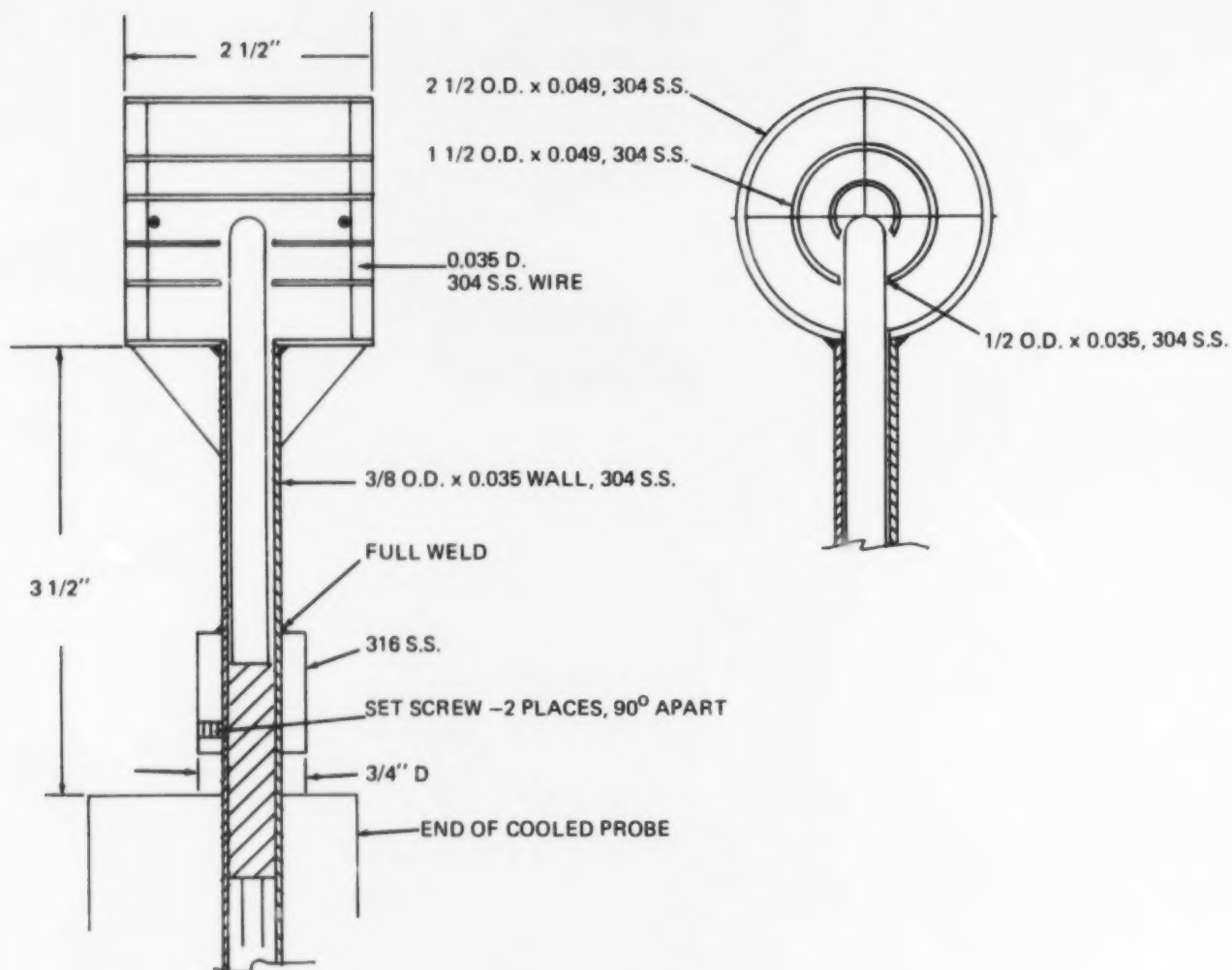
*Courtesy of Mr. James Dorsey, Control Systems Laboratory.

SECTION AA REFERS TO ORIGINAL PROBE ARRANGEMENT
 SECTION BB REFERS TO MODIFICATION OF PROBE WHERE,
 AT SECTION AA, PROBE WAS CUT SUCH THAT SAMPLE
 WAS PULLED STRAIGHT THROUGH, RESULTING IN
 PROBE BEING 2 INCHES SHORTER.



1. THERMOCOUPLE WELL - 3/8-INCH BWG20, 316 S.S.
2. COOLING WATER - 1/4-INCH 1 TYPE L, COPPER TUBING
3. SAMPLE OUTLET - PYREX GLASS LINER, 5/8-INCH O.D.
16 M.M., WITH 28/12 BALL JOINT
4. SAMPLE TUBE RECEIVER - 1/2-INCH SCHEDULE 10, 316 ELC-S.S.
5. SAMPLE INLET - 3/8-INCH SCHEDULE 80, 316 ELC-S.S.
6. S-TYPE PITOT TUBE - 3/8-INCH BWG 20, 316 S.S.
7. COOLING WATER INLET - 1/4-INCH SCHEDULE 40, 316 ELC-S.S.
8. COOLING WATER OUTLET - 1/4-INCH SCHEDULE 40, 316 ELC-S.S.
9. PROBE JACKET END PLATE - 316 ELC-S.S., 11 GAGE 2B FINISH
10. PROBE JACKET - 1 1/4-INCH SCHEDULE 10, 316 ELC-S.S.
11. MOISTURE SAMPLING TUBE - 1/4-INCH BWG20, 310 S.S.
12. WORM-DRIVEN HOSE CLAMPS

Figure 18 Water-cooled sampling probe.



MODIFICATIONS

2 1/2 O.D. WALL INCREASED TO 0.065
 1 1/2 O.D. WALL INCREASED TO 0.065
 3/8 O.D. TUBE REPLACED WITH 1/4 INCH SCH 40, 316 S.S.
 1/2 O.D. TUBE OMITTED
 0.035 INCH WIRE REPLACED WITH 0.093 316 ELC S.S. ROD
 WEDGE WAS ADDED FOR STRUCTURAL SUPPORT, 2 x 1 90° TRIANGLE
 MATERIAL - 316 ELC S.S., 11 GAGE

Figure 19 Thermocouple radiation shield

Instrumentation

The following instrumentation was used for monitoring the concentration of the species indicated:

H₂ — Carle Model 8004 gas chromatograph with automatic timer and sample injection.

CO₂, O₂ — Carle Model 8004 gas chromatograph with automatic timer, sample, and N₂, CO column switch.

CO₂ — Beckman Model 315B NDIR.

— MSA Model 300 LIRA NDIR.

CO — MSA Model 300 LIRA NDIR (two).

HC — Beckman Model 108A FID.

O₂ — Beckman Model 742 polarograph.

Pressure — MKS Baratron capacitance gauge.

Temperature — L and N multipoint recorder.

The main gas-probe sampling pumps were made by Gast (Models 0322), and the analytical measurement pumps were made by Metal Bellows Corp. (MB-21). Rockwell dry gas meters were also used, and RAC inline aluminum filters were used for both locations. All fittings were Swagelok, and Nupro needle and toggle valves were used. All components of the gas analysis system were brass and copper.

Photographs of Equipment

Figures 20 through 23 are photographs of the various pieces of equipment in the analytical shed. Figure 20 is the three main gas sampling systems as represented earlier in Figure 15. Figure 21 shows the multipoint temperature recorder and the original three-pitot pressure-measurement system. The pressure system was later expanded to the 10-point measurement manifold. Figure 22 is a view including the NDIR CO and FID HC analyzers and several of the recorders. Figure 23 shows the two original gas chromatograph analyzers (one was later replaced with continuous CO₂ and O₂ analyzers) and the CO NDIR analyzers.

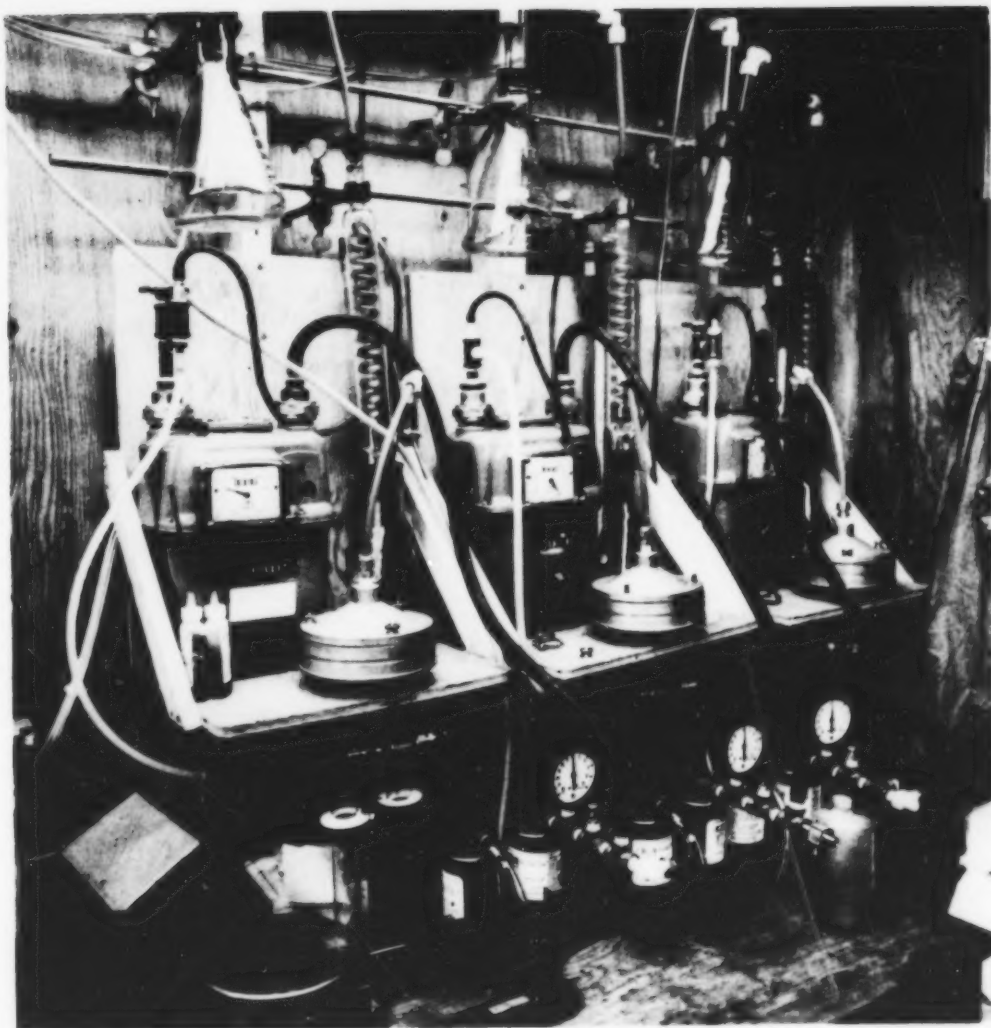


Figure 20 Gas sampling lines, pumps and meters.

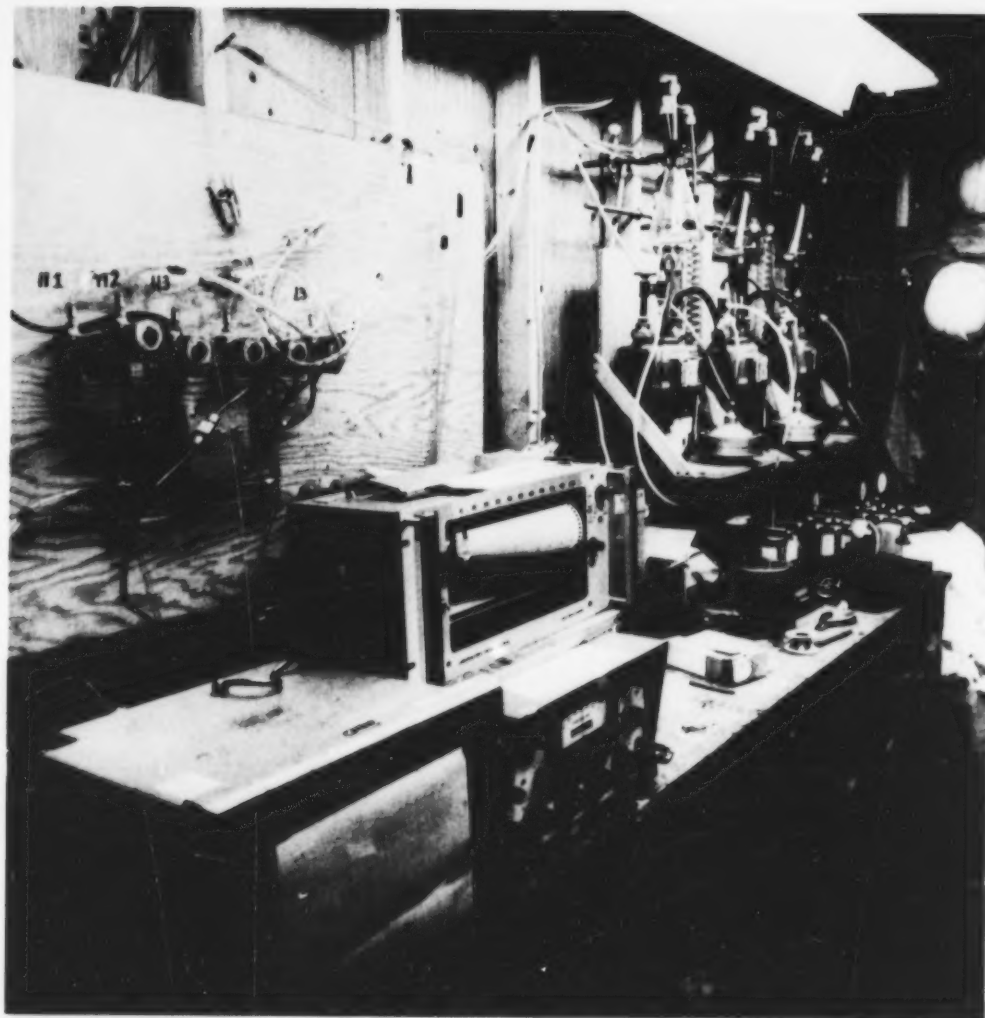


Figure 21 Recording temperature and pressure systems.

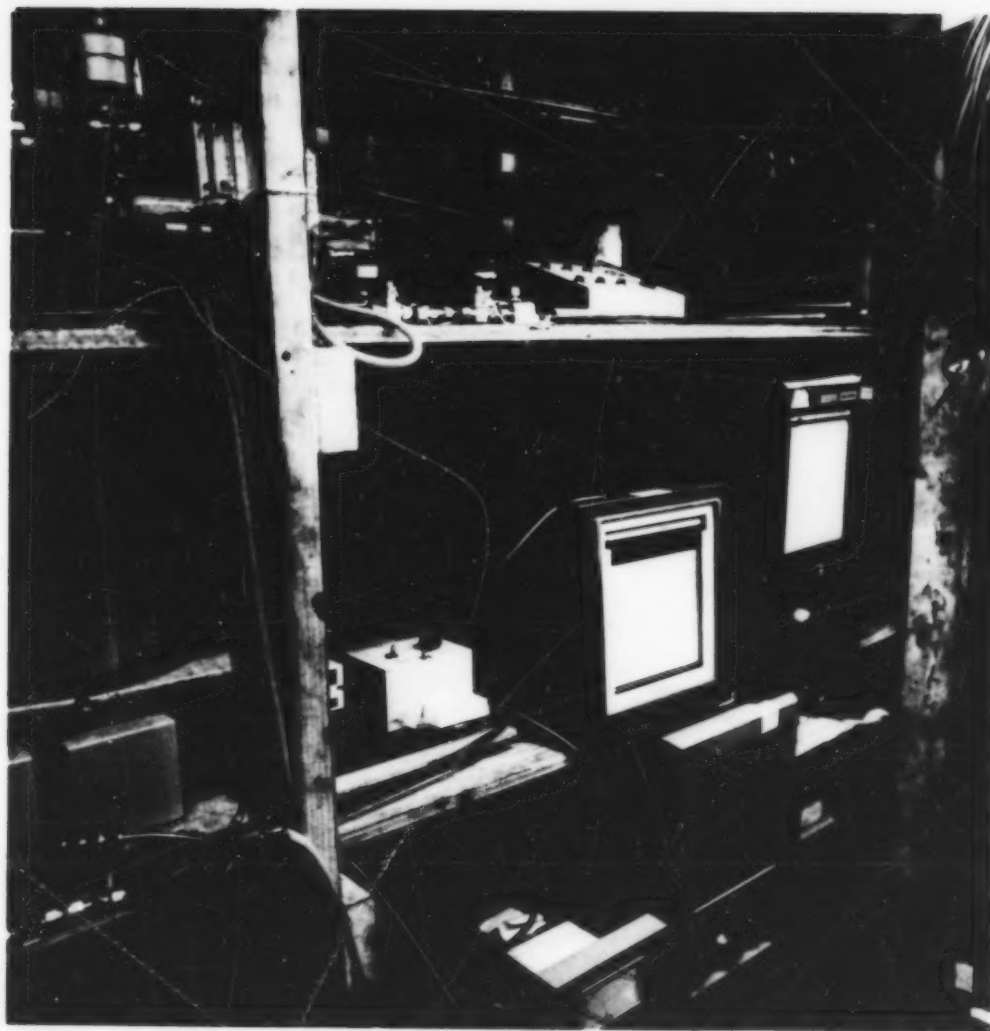


Figure 22 NDIR and FID analyzers and recorders.



Figure 23 Gas chromatograph and NDIR analyzers.

DATA REDUCTION

The data reduction was primarily a three-step operation. The form of the data was first changed from percent displacement on a strip chart recorder to a tabular listing of data in appropriate units. In the second step, the data for a particular run were averaged so that each run could be characterized by a simple set of output parameters. Finally, the validity of the output parameters was checked by mass and energy balances.

Data Recording and Averaging

The first and second steps were purely mechanical manipulations of the recorded data. In reading data points from a continuous strip chart recorder, five-minute averages were taken at five-minute intervals. This particular method was chosen because of the five-minute sample time for each of the breech zones. The raw averaged data were then applied to the appropriate calibration curve to establish the data in desired units of temperature, concentration, or velocity. The moisture content data were recorded in 15-minute intervals as 15-minute averages; this timing corresponded to one complete cycle of the breech zone sampling system. To calculate the set of output parameters for each run, the tabulated data were time-averaged for the entire run. The output parameters are listed below:

- Temperature
Breech zones 1, 2, and 3 and the average temperature over the entire breech
- Flow rate
Velocity in breech zones 1, 2, and 3
Total molar flow rate
- Composition of exhaust gas
Breech zones 1, 2, and 3 – CO_2 , CO , CH_4 , H_2O , and O_2 , and the average molecular weight
For total breech – CO , CH_4 , O_2 , N_2 , CO_2 , and H_2O
- Heat output of system

These output parameters were used for the remainder of the analysis.

Data Checks – Mass and Energy Balances

The difficulties encountered in obtaining valid data from an operating municipal incinerator are well known. The conditions under which sampling must take place are hot (1200 to 2000°F), dirty (molten fly ash, heavy organics, high moisture content), and not chemically constant from second to second or, on the average, from hour to hour. These difficulties do not preclude the possibility of obtaining valid data, but they do place increased significance on material and energy balances as a means of verifying the internal

consistency of the data. Although we accept that mass and energy balances are not the necessary and sufficient criteria for proving the validity of data, reasonably good balances do tend to increase the credibility of the data and, hence, our confidence in it.

As a rule, then, whenever balances can be closed, the data are considered valid. For experiments in which the balances did not close, the data were rejected. In the latter cases, we attempted to explain the most probable source of error or inconsistency.

The Mass Balances

The overall mass balance over the incinerator was impossible to obtain because a large fraction of the air entering the furnace was not metered (leakage). The amount of leakage was determined by the difference between the exiting gases and the forced air and combustion-generated gases.

Carbon — The only important source of carbon in the system is the refuse feed. It is assumed that the average refuse at Newton is 50% combustible, and that these combustibles have an empirical formula of $\text{CH}_{1.585}\text{O}_{0.675}$ based upon previous data for Newton [8]. Carbon leaves the system as CO_2 , CO, various hydrocarbons, soot, and unburned material which falls out of the incinerator as ash. For a typical run the distribution of carbon among these various forms is shown in Table 4.

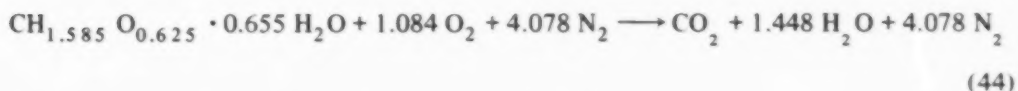
Table 4. CARBON DISTRIBUTION PROFILE

Component	Pounds (mols C/hr)	Percent
<u>Carbon In</u>	415.71	100
Refuse	413.35	99.43
CO_2 in air	2.36	0.57
<u>Carbon Out</u>	415.71	100
CO_2	410.31	98.46
CO	1.49	0.36
Hydrocarbons	0.56	0.13
Soot	0.21*	0.05
Ash	4.14*	1.00

* Arthur D. Little, Inc., estimates.

Because carbon dioxide represents such a large percentage of the total "carbon out," in the remainder of the analysis we have ignored the contributions from CO, hydrocarbons, soot, and the ash. By the same reasoning only the "carbon in" as refuse will be utilized in further carbon balances.

For these specific data, the combustion of refuse is represented by the following formula:



This relationship suggests that 1 pound mol of CO_2 should be produced for every 23.585 pounds of combustible (or 47.17 pounds of refuse). Because of the uncertainty and variations in the bulk density of refuse, as well as the voids which are present in the feed hopper, it is difficult to calculate the exact feed rate. However, for the purposes of the material balances, feed rates were calculated assuming a density of 13 lb/ft³ and that there were no voids in the feed hopper. It can be seen from Table 4 that the carbon in and the carbon out are in reasonable agreement. From the information in this table one can calculate that the average amount of combustible fed into the furnace was 51.5% of the refuse ($\rho = 13 \text{ lb/ft}^3$). For the purposes of the current work, this agreement is adequate to verify that the techniques and apparatus are appropriate.

Hydrogen — Hydrogen enters the system as water vapor in the air, as moisture in the refuse, as bound hydrogen in the refuse, and as water evaporated from the ash quench system. The hydrogen leaves the system as water vapor and hydrocarbons in the exhaust gases and as moisture and bound hydrogen in the unburned combustibles. A hydrogen balance for a typical run is shown in Table 5. As in the case of the carbon balance, the small contributors have been ignored; that is, hydrocarbons and hydrogen in ash either bound or as moisture and the water vapor in the air from the ash quench system. The "hydrogen out" is all measured as water vapor. The "hydrogen in" is calculated from the "carbon dioxide out" (which has previously been related to amount of refuse fed). The empirical relationship states that the mols of water released is equal to 1.448 times the mols of CO_2 released. As can be seen, the water balance (which implies the hydrogen balance) is good enough to justify the use of the data. In the current study we found the average moisture content of refuse to be 24.6%, as compared to the assumed value of 25%.

Table 5. HYDROGEN BALANCE FOR TYPICAL RUN

Component	Pound (mols H/hr)	Percent
<u>Hydrogen In</u>	1373.24	100
H as H ₂ O in Air	122.90	8.94
H as H ₂ O in Refuse	541.58	39.45
H (bound) in Refuse	655.16	47.71
H as H ₂ O from Ash Quench*	53.60	3.9
<u>Hydrogen Out</u>	1373.24	100
H ₂ O in Gases	1359.13	98.97
H as Hydrocarbons	2.13	0.16
H ₂ O in Ash	5.42	0.39
H (bound) in Ash	6.56	0.48

*Arthur D. Little, Inc., estimate.

Oxygen — The oxygen in the system comes from the atmospheric oxygen and water vapor, the moisture in the refuse, the water evaporated from the ash quench, and the bound oxygen in the refuse combustibles. The oxygen leaves the incinerator as atmospheric oxygen, carbon dioxide, carbon monoxide, and water vapor. A typical run has an oxygen balance similar to that presented in Table 6.

Table 6. OXYGEN BALANCE IN TYPICAL RUN

Component	Pound (mols O/hr)	Percent
<u>Oxygen In</u>	3673.65	100
O as O ₂ in Air	3056.27	83.20
O as H ₂ O in Air	61.45	1.67
O as H ₂ O in Refuse	270.79	7.37
O as H ₂ O from Ash Quench*	26.80	0.73
O as Bound Oxygen in Refuse	258.34	7.03
<u>Oxygen Out</u>	3673.65	100
O as O ₂ in Gases	2166.68	58.98
O as H ₂ O in Gases	679.56	18.50
O as CO ₂ in Gases	820.62	22.34
O as CO in Gases	1.49	0.04
O as H ₂ O in Ash*	2.71	0.07
O as Bound Oxygen in Ash*	2.59	0.07

*Arthur D. Little, Inc., estimate.

The small contributors — CO, oxygen in ash as moisture and bound water, and water from ash quench — are not considered when solving the oxygen material balance.

A summary of the average oxygen balances is given in Table 6. As can be seen, these balances are adequate for the current work.

The Energy Balance

The energy balance over the incinerator has been used in the same way as the mass balances to verify the experimental techniques and the analytical apparatus. Essentially, there are two ways of conducting the energy balance. The first is to assume a constant heating value for refuse and check the heat actually liberated with the theoretical heat available. The other is to compare the heat liberated per unit of combustion product formed. The latter is the chosen course because it does not presume knowledge of the heat content of the raw refuse feed or the feed rate.

It is known that about 5650 Btu's are released per pound mol of CO_2 liberated in the combustion of many organic materials. Therefore, one could conduct the heat balance by knowing the mass flow, composition, temperature, and heat capacity of the gas stream. On the other hand, the same heat balance could be accomplished knowing only the fraction of the flue gas which is CO_2 and the temperature. If we assume that the heat capacity of the flue gas stream does not change with composition, then the temperature of the flue gases will be a linear function of the fraction of CO_2 (Appendix D, Eq. D-31). Furthermore, the theoretical line passes through the adiabatic flame temperature at the stoichiometric CO_2 concentration and through ambient temperature at a CO_2 concentration equal to zero. Figure 24 shows the data on such a plot of temperature vs CO_2 concentration. Because the system is not adiabatic, it is reasonable to assume that the dotted curve represents the kind of deviation from the theoretical that we might expect, that is, the largest deviation at higher temperatures and no deviation at ambient. Rotation of the line about the average temperature would also be expected as a result of radiation. From the slope of a line through the plotted points, which is about $.004\% \text{ CO}_2 / ^\circ\text{F}$ (or $260^\circ\text{F} / \%\text{CO}_2$), we have calculated that the heating value of the refuse is about 4650 Btu/lb based upon Eq. (D-31) of Appendix D.

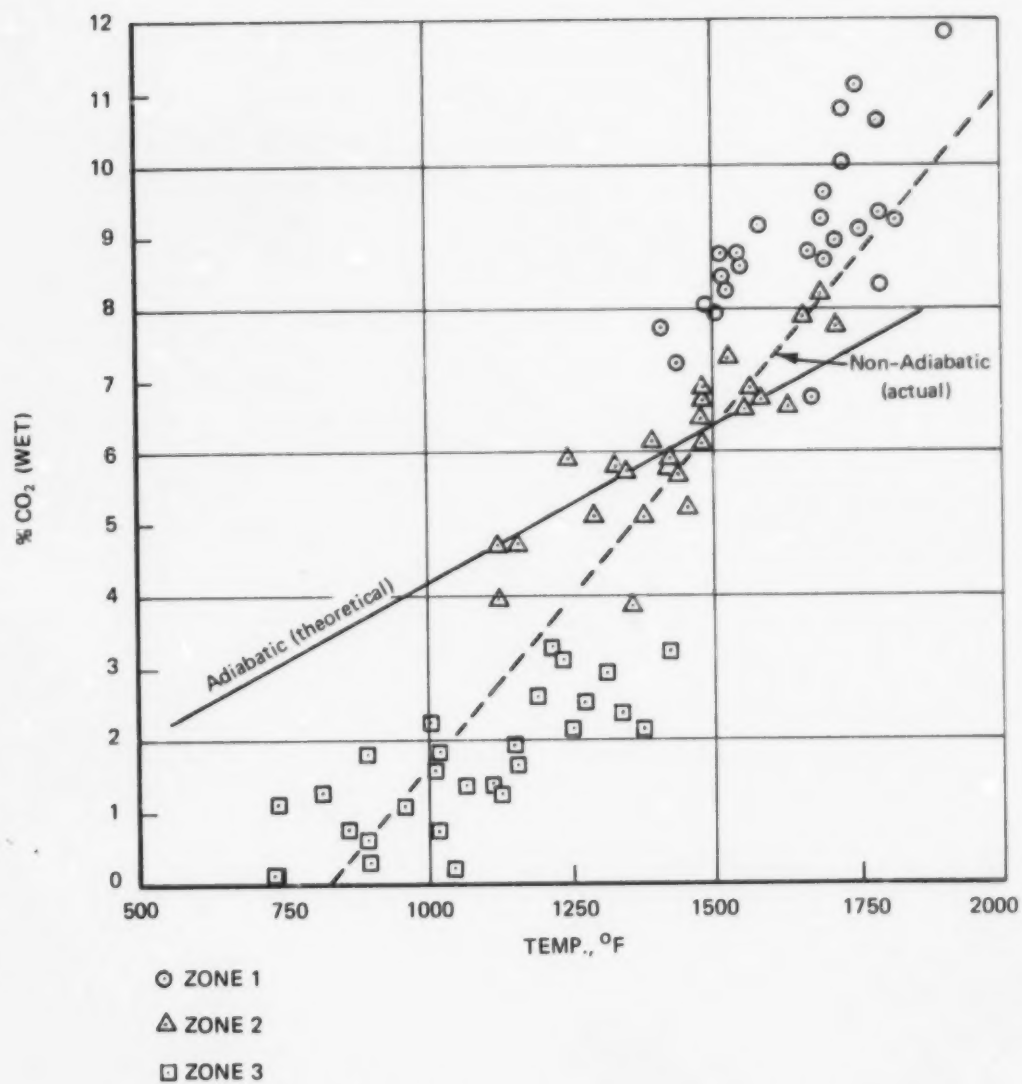


Figure 24 CO₂ vs. temperature.

SECTION VII

TEST RESULTS

The strategy we espoused for controlling combustible emissions in Section V was predicated on two separate mixing concepts:

- Macroscale mixing to achieve uniformity throughout the furnace; and
- Microscale mixing to promote burnout of combustibles at the molecular level.

The equations that we presented as the basis for each of the two mixing models are verified in this section. The data were taken at an existing municipal incinerator which was retrofitted with overfire air jets for this program.

The test equipment and procedures were discussed in Section VI. In this Section, we will describe the specific tests that we carried out and discuss the results of those tests. A compilation of the test data is given in Appendix A.

MACROSCALE MIXING MODEL

Overfire jets can be used to achieve uniform mixing throughout the furnace. The macroscale model proposes a design basis for overfire air jets based upon:

- Ivanov's equation for jet penetration, and
- Use of Bernoulli's equation to characterize the flow patterns within the furnace.

The method for calculating the penetration of an overfired jet into the furnace was suggested by Ivanov and has been reviewed in Section V. The design equation (Eq. (3)) is repeated below:

$$L_p = 1.6 \sqrt{\frac{\rho_j}{\rho_c}} \frac{d_j u_j}{u_c} \quad (3)$$

The equation is dependent upon the jet diameter, its velocity, and the crossflow velocity of gas in the furnace. In a series of jet tests the penetration distance was estimated visually and was compared with the calculated value for penetration. Figure 25 shows a plot of penetration vs jet velocity for 5-inch diameter jets firing into a crossflow velocity field of 15 ft/sec. The crossflow velocity was calculated using the fluid-flow equations discussed previously. Clearly, the Ivanov relation is adequate for estimating jet penetration, and the deviation from the theoretical relationship is well within experimental error.

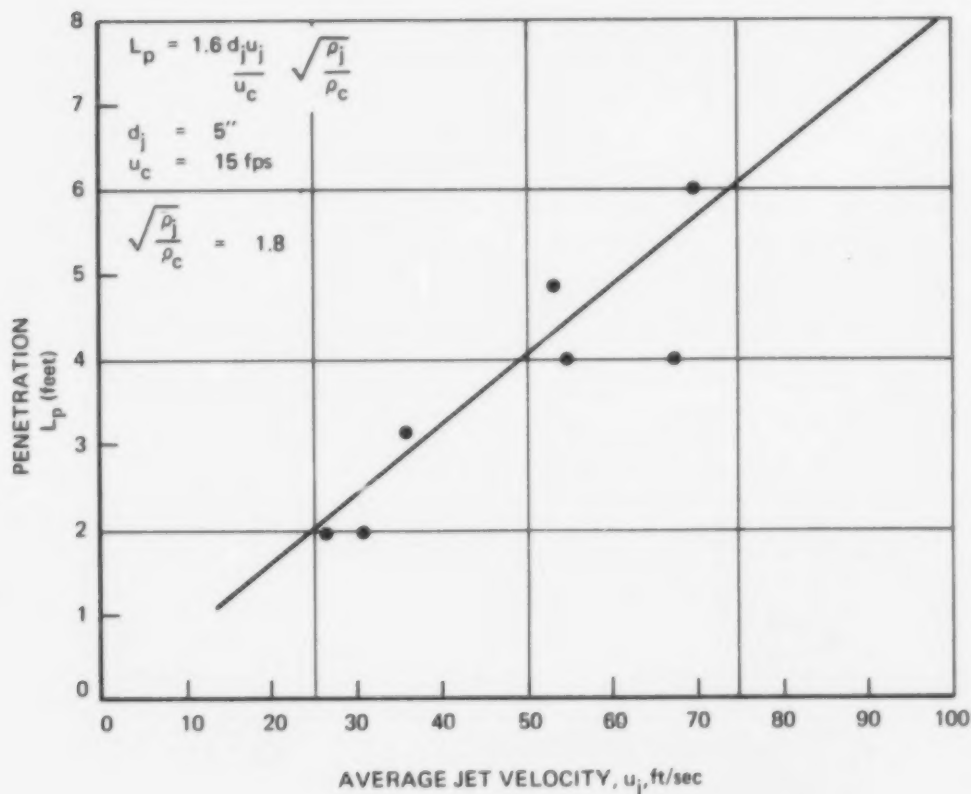


Figure 25 Penetration vs. jet velocity.

FLUID-FLOW PATTERNS

The fluid-flow model developed in the mixing study is based on the premise that Bernoulli's equation can be applied along the streamline of flow in an incinerator so that both the residence time and the crossflow velocity at the level of the jets can be calculated.

During the tests, velocity and temperature were measured in the breach of the incinerator. The profiles clearly showed the stability of the hot zone at the top of the breach, but the data did not show large difference between the "hot" and "cold" zone velocities. The equations derived for a sealed furnace [Eq. (8)] could not correlate the data, as witnessed by the random appearance of a plot of u^2 vs $2g(z-z_o) [(T_H/T_C) - 1]$. On the other hand, Eq. (9) derived for unsealed furnaces could correlate the data reasonably well. A plot of u^2 vs $[(T/T_o) - 1]$ where T_o is the ambient temperature and T is the

temperature of either the hot zone or cold zone is shown in Figure 26. The plot should be linear if Eq. (9) repeated here is applicable to incinerators:

$$u_H^2 = u_{H_0}^2 + 2g(z_H - z_0) \left(\frac{T_H}{T_0} - 1 \right) \quad u_C^2 = u_{C_0}^2 + 2g(z_C - z_0) \left(\frac{T_C}{T_0} - 1 \right) \quad (9)$$

With the exception of a few spurious points, the agreement with the theoretical relationship is excellent. The solid line in the figure represents the best fit through the data points corresponding to a furnace height of 12 feet and an initial velocity of 18.4 ft/sec. The dashed line represents the curve passing through the origin, so that the furnace height is 13.7 feet, while the initial velocity is zero. Because of the gross nature of the flow patterns within the furnace, the analytical distinction between the height of 12 feet and 13.7 feet is not possible. Either curve could be justified as confirmation of the Bernoulli approach; neither is so well defined that it may be used to predict initial velocities.

Two conclusions can be drawn from this analysis:

- 1) Bernoulli's equation can, indeed, be used for design purposes in determining exit velocities and approximate furnace residence times; and
- 2) In a furnace which has considerable air leakage, both the hot and cold zones will accelerate according to the Bernoulli relation. The driving force for the acceleration is the buoyancy of gases within the furnace as compared to outside the furnace.

Verification of Eq. (10) of Section V, which describes the difference in velocity of the hot zone vs the cold zone, was not possible. The accuracy of the data does not warrant such fine tuning of the fluid-flow model.

MICROSCALE MIXING MODEL

The theories presented to characterize the microscale mixing introduced the use of Hottel's unmixedness factor and Corrsin's equation as fundamental relationships characterizing the three T's on a microscopic scale. Hottel's and Corrsin's concepts have been verified in controlled situations, but because of the large variations in incineration data it is unrealistic to hope for conclusive evidence that will provide exact quantitative relations. The data are sufficiently accurate, however, to demonstrate the consistency of the data with the two concepts.

The Unmixedness Factor

In Section V, the relationship between the unmixedness factor suggested by Hottel (4) and the ratio Φ of available oxygen to oxygen required to complete combustion was

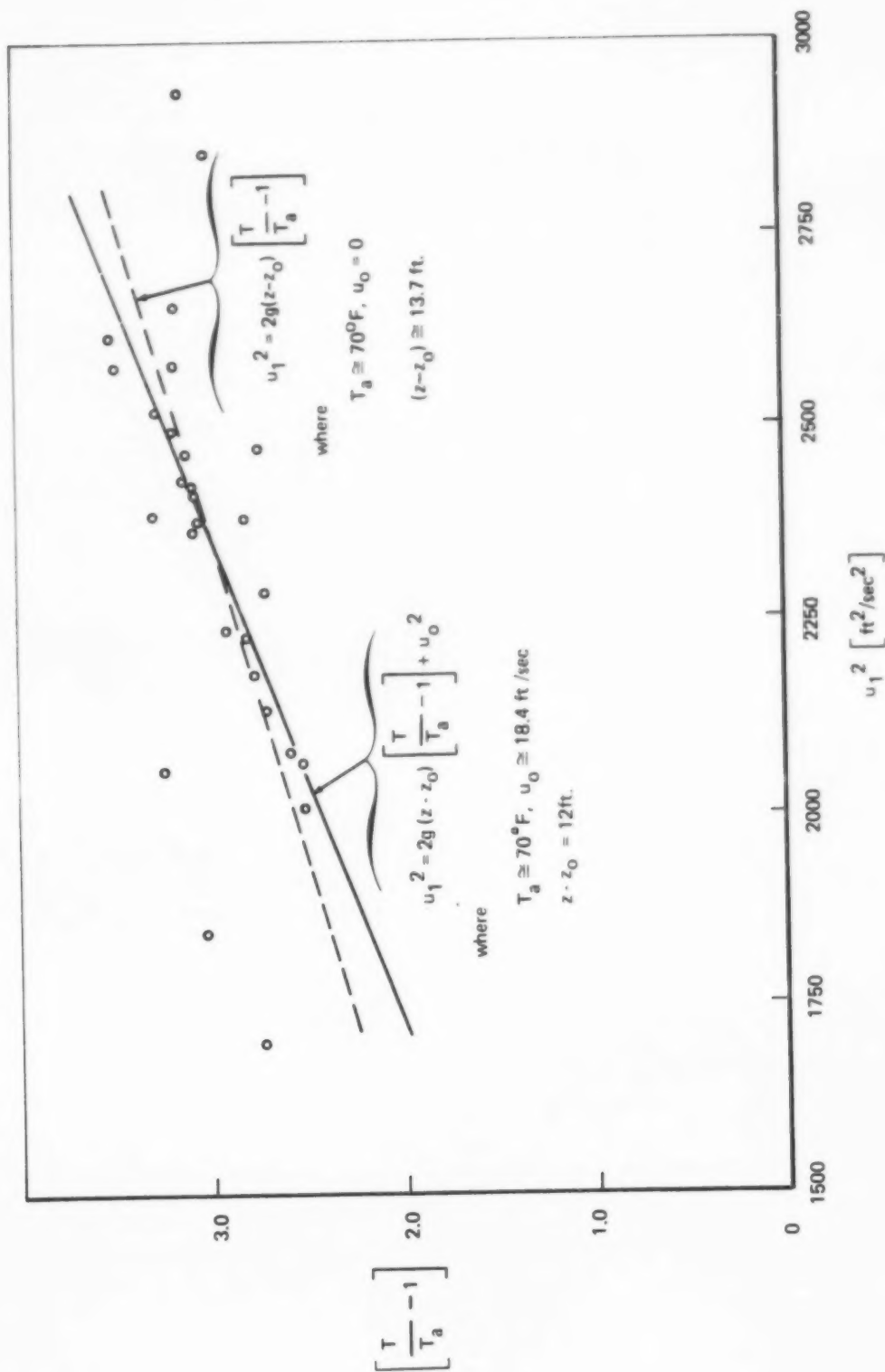


Figure 26 Confirmation of Bernoulli's equation.

presented. The relation is derived in Appendix F. The procedure for calculating the value of the unmixedness factor is outlined in Appendix B. Supporting data required to characterize the refuse bed offgas is presented in Appendix E.

The unmixedness factor can be calculated for each gas sample that is taken. Its value is dependent only upon the gas analyses and the empirically determined value for CO/CO₂. The experimental value for $\sqrt{c'^2}/\Delta y$ is determined from Figure 3, based on the value of Φ [(available oxygen)/(oxygen required to complete combustion)], determined from the gas analysis. The value of the unmixedness factor, $\sqrt{c'^2}$, is calculated from the following equation which is based upon Eq. (B-12):

$$\sqrt{c'^2} = \left(\frac{\sqrt{c'^2}}{\Delta y} \right) \left[y_s \left(1 - \frac{\%CO_2}{\%CO_2|_s} \right) \right] \quad (45)$$

In Appendix B, the quantity y_s is shown to be equal to 0.78 when CO/CO₂ is equal to 0.5. The composition of the combustibles and combustion products (without entrained air) is also given.

In Table 7, the experimental value for $\sqrt{c'^2}$ is given for each breach zone and each of the 28 jet tests.

To compare the experimental value for unmixedness factor with the theoretical one, the following data are required:

- original CO/CO₂ off the bed, and
- relationship between temperature and CO₂ in the breach.

In these calculations, a value of 0.5 for CO/CO₂ has been assumed.

The plot of temperature vs %CO₂|_w was given previously in Figure 24.

The data can be represented by the linear relationship:

$$\frac{\%CO_2|_w}{10\%} = (\text{Temp} - 500^\circ\text{F})/1500^\circ\text{F} \quad (46)$$

The calculation procedure is as follows:

1. For a given temperature, calculate %CO₂|_w from the equation given above;
2. Calculate $\Delta y = y_s \left(1 - \frac{\%CO_2|_w}{15.1\%} \right)$;
3. For a given value of $\sqrt{c'^2}$, calculate $\left(\sqrt{c'^2}/\Delta y \right)$;

Table 7. SUMMARY OF UNMIXEDNESS FACTORS

Zone I					Zone II					Zone III					
Run #	Emission Factor	Temp.	$\sqrt{c^{-1}}_{\text{expt}}$	$\sqrt{c^{-1}}_{\text{theo}}$	Δc^{-1}	Emission Factor	Temp.	$\sqrt{c^{-1}}_{\text{expt}}$	$\sqrt{c^{-1}}_{\text{theo}}$	Δc^{-1}	Emission Factor	Temp.	$\sqrt{c^{-1}}_{\text{expt}}$	$\sqrt{c^{-1}}_{\text{theo}}$	Δc^{-1}
22.1	1	1721	<0.12	0.62	-	1015	1446	0.222	0.237	-0.015	1314	1029	0.270	0.280	-0.010
22.2	1	1508	<0.14	0.76	-	583	1387	0.210	0.223	-0.013	274	1004	0.275	0.240	-0.035
22.3	1	1700	<0.12	0.65	-	244	1458	0.180	0.172	0.008	931	1100	0.250	0.278	-0.028
23.1	127	1715	0.145	0.116	0.029	677	1446	0.210	0.223	-0.013	480	1046	0.235	0.239	-0.004
23.2	87	1437	0.165	0.157	0.008	909	1125	0.250	0.244	0.006	979	737	0.280	0.241	0.039
23.3	330	1717	0.160	0.149	0.011	3705	1350	0.290	0.312	-0.022	12 162	871	0.250	0.350	-0.100
23.4	252	1421	0.190	0.200	-0.010	2707	1158	0.290	0.287	0.003	785.3	896	0.370	0.338	0.032
24.1	144	1647	0.150	0.174	-0.024	535	1385	0.210	0.210	0	1102	892	0.270	0.262	0.008
24.2	60	1694	0.135	0.115	0.020	18	1470	0.150	0.149	0.001	1633	1152	0.270	0.296	-0.026
24.3	55	1551	0.150	0.126	0.024	24	1341	0.160	0.159	0.001	2461	949	0.300	0.308	-0.006
24.4	409	1730	0.165	0.132	0.033	24	1426	0.150	0.160	-0.010	2508	1071	0.295	0.308	-0.013
25.1	652	1900	0.155	0.099	0.056	499	1696	0.170	0.167	0.003	823	1283	0.230	0.304	-0.074
25.2	976	1796	0.180	0.193	-0.013	462	1527	0.185	0.181	0.004	1242	1225	0.250	0.270	-0.020
25.3	106	1400	0.170	0.147	0.023	508	1248	0.225	0.219	0.006	2080	875	0.295	0.297	-0.002
25.4	91	1523	0.160	0.135	0.025	1387	1131	0.265	0.265	0	5141	742	0.360	0.375	-0.015
31.1	157	1750	0.145	0.093	0.052	267	1543	0.175	0.173	0.002	823	1193	0.240	0.265	-0.025
31.2	46	1507	0.170	0.135	0.035	442	1261	0.215	0.221	-0.006	193	836	0.230	0.241	-0.011
31.3	26	1521	0.145	0.125	0.020	572	1308	0.220	0.214	0.006	142	1025	0.210	0.231	-0.021
32.1	243	1825	0.140	0.139	0.001	40	1707	0.130	0.138	-0.008	1403	1381	0.240	0.290	-0.050
32.2	207	1683	0.155	0.130	0.025	187	1657	0.165	0.187	-0.022	84	1408	0.170	0.211	-0.041
32.3	235	1698	0.155	0.161	-0.006	416	1609	0.180	0.198	-0.018	1042	1344	0.230	0.272	-0.042
33.1	275	1777	0.150	0.113	0.037	550	1545	0.190	0.202	-0.012	2449	1251	0.280	0.304	-0.024
33.2	86	1666	0.140	0.130	0.010	416	1485	0.190	0.193	-0.003	2012	1240	0.270	0.283	-0.013
33.3	505	1753	0.150	0.159	-0.009	545	1585	0.190	0.195	-0.005	3340	1306	0.290	0.311	-0.021
34.1	272	1793	0.150	0.138	0.012	320	1478	0.185	0.185	0	2484	1150	0.290	0.272	0.018
34.2	67	1554	0.150	0.129	0.021	502	1417	0.205	0.214	-0.009	2594	1108	0.295	0.312	-0.017
34.3	61	1497	0.155	0.138	0.017	596	1367	0.215	0.230	-0.015	2852	1023	0.305	0.315	-0.010
34.4	54	1597	0.140	0.118	0.022	412	1481	0.190	0.195	-0.005	3689	1115	0.320	0.325	-0.005

4. Determine Φ from Figure 3;
5. Calculate %O₂ from the material balance equation (Eq. D-26 of Appendix D):

$$\frac{\%O_2}{21\%} + \frac{\%CO_2|_w}{15.1\%} = 1 ;$$

6. Calculate O₂ required, according to the relation:

$$O_2 \text{ req} = (\%O_2) / \Phi ;$$

7. Calculate the emission factor according to the equation:

$$CO \text{ (ppm)} | 12\% = (O_2 \text{ req}) \frac{CO|_1}{n_{O_2}} \left(\frac{12\%}{\%CO_2} \right)$$

Note that for refuse:

$$\frac{\%CO_2|_w}{15.1\%} = \frac{\%CO_2|_d}{19.5\%} \quad (47)$$

Using the above steps, we calculated the data shown in Figures 9, 10, and 11. The procedure shows that emission factors can be calculated if the temperature and unmixedness factor are specified and if the original composition of the offgas is known.

The theoretical value for $\sqrt{c^{T2}}$ has been determined from Figure 8 for each of the jet tests, and these values are tabulated in Table 7.

A careful examination of the data reveals that in all cases, even in the quench zone, high emissions are reflected in high unmixedness factors. The agreement between experimental and theoretical values overall is 0.003 or within 1% of the expected value. The agreement is best in Zone II where the data from Figure 24 are best fitted by Eq. (46).

The parameter $(CO/CO_2)_0$ has no effect on the relative values of theoretical and experimental unmixedness factors, since the ratio was used to calculate both. However, a change in the value of $(CO/CO_2)_0$ results in a change in the absolute value of $\sqrt{c^{T2}}$. Hence, to attempt to place any weight on the actual numerical values in Table 7 would be stretching the validity of the data. With respect to the validity of the concept, the correlation of the data is remarkably well fitted to the unmixedness factor concept. Without this characterization factor, the more sophisticated advances in the theoretical understanding of incinerator mixing do not readily follow.

Corrsin's Equation

The verification of Corrsin's Equation could be based upon the defining equation, (Eq. (21)):

$$\left(\sqrt{c^{T2}}\right) = \left(\sqrt{c^{T2}}\right)_0 e^{-\beta t}, \quad (21)$$

or alternatively upon the comparative equation (Eq. (40)):

$$\left(\sqrt{c^{T2}}\right)_j = \left(\sqrt{c^{T2}}\right)_b \left[1 + \frac{y_s}{\Delta y_b} \left(\frac{\%CO_2|_b}{\%CO_2|_s} \right) (1-x) \right] \left[\frac{f(\phi)_{j0}}{f(\phi)_{b0}} \right] e^{-[(\beta t)_j - (\beta t)_b]} \quad (40)$$

In Appendix E, tests are described wherein offgas compositions were measured directly over the refuse bed. From these data the initial value of the unmixedness factor $\sqrt{c^{T2}}$ was calculated. The calculated values for the two complete bed tests are shown in Table 8. The scatter in the data was so great that no meaningful conclusions were possible.

Table 8. SUMMARY OF TEST RESULTS OVER BED⁽¹⁾

	Sample Sort			
	1	2	3	Average
Bed Test 33				
CO ₂	5.93	1.31	5.8	4.35
O ₂	13.4	18.9	14.6	15.63
CO	3.2	0.26	0.13	1.20
H ₂	1.72	0.21	0.09	0.67
HC	1.16	0.01	0.03	0.40
H ₂ O ⁽²⁾	11.51	2.15	8.84	7.50
H ₂ O/CO ₂	1.94	1.64	1.52	1.72
CO/CO ₂	0.54	0.20	0.02	0.28
K _{wg}	3.61	2.03	2.2	3.1
Temperature	1835	1905	1928	1889
Bed Test 2				
CO ₂	13.3	6.4	12.0	10.6
O ₂	6.0	14.2	8.6	9.6
CO	3.35	1.00	0.26	1.54
H ₂	2.7	0.77	0.19	1.22
HC	1.94	0.34	0.39	0.89
H ₂ O	12.9	19.4	26.4	17.57
H ₂ O/CO ₂	0.97	3.03	1.70	1.66
CO/CO ₂	0.25	0.16	0.02	0.15
K _{wg}	1.20	3.94	2.33	2.09
Temperature	1787	1758	1811	1785

1. Complete sets of data are limited due to experimental difficulties directly over the bed.

2. Calculated from material balance.

In pursuing the alternative approach to verification of the Corrsin relation, note that neither the parameter x nor the parameter $(\beta t)_b$ are well known or clearly defined. The value for x could be defined by Eq. (28) in Section V, based upon the ratio $\%CO_2|_j/\%CO_2|_b$. One might also define x based on temperature by substituting Eq. (46) into Eq. (28) of Section V to yield:

$$(1-x) = 1 - \frac{(T_j - 500)}{(T_b - 500)} \quad (48)$$

Finally, one might also define x based upon flow rates (Eq. (30)). For example, the different calculated values for x in each of the test conditions of Test 31 are shown in Table 9 to be in relatively good agreement. The values for Test 33 are grossly apart.

Table 9. COMPARISON OF ESTIMATES FOR $(1-x)$

Test No.	Basis for $(1-x)$		
	Eq. (28)	Eq. (47)	Eq. (30)
31-2	0.284	0.194	0.290
31-3	0.260	0.180	0.294
33-1	0.163	0.087	0.277
33-3	0.139	0.019	0.256

The reasons for the discrepancies in the above values for x are:

- Errors in analytical data,
- Changes in total flow rates into the furnace, and
- Changes in composition or heating value of refuse.

To determine the effect that different values for x might have on the resulting calculations, consider the calculations summarized in Table 10. In these calculations, based upon Tests 31-1 (base condition) and 31-2 (jet condition), x is chosen arbitrarily and Δy is defined by the equation:

$$\Delta y = y_s \left(1 - \frac{\%CO_2|_b}{\%CO_2|_s} \cdot x \right) \quad (49)$$

Note that as the value for $(1-x)$ decreases, the value for the experimentally determined unmixedness factor $\sqrt{c'^2}$ also decreases (due to the change in the calculated value for Δy , according to Eq. (49) above), and the predicted value for $\sqrt{c'^2}$ increases, primarily due to the decrease in the value of $(\beta t)_j$. The result is the rapid convergence of the experimental and predicted values for unmixedness factor to the extent that agreement with the Corrsin relation can be shown.

In the first set of calculations βt was assumed to be equal to 0.25. On the other hand, the value of (βt) is not clearly defined, because it depends on the arbitrary

Table 10. CALCULATION RESULTS USING DIFFERENT VALUES FOR α

Test	(1-x)	From Data (Appendix B)			From Theory			Application of Eq. (41)		$f(\psi_0)/f(\psi_{b0})$	βt	$\sqrt{c^2}$	$\Delta\%$
		Φ	$f(\Phi)$	$\Delta y(x)$	$\sqrt{c^2}$	ψ_0	$f(\psi_0)$	$\sqrt{c^2}_0$	$\frac{Y_1}{1 + \Delta Y} \left(\frac{\%CO_2/b}{\%CO_2/s} \right) (1-x)$				
31-1	0	445	0.455	0.207	0.094	86	0.584	0.121	1.0	1.0	0.250	0.094	-
31-2	0.284	3339	0.370	0.371	0.137	203	0.500	0.186	1.791	0.856	0.745	0.088	-35.8
	0.150	3339	0.370	0.294	0.109	138	0.535	0.157	1.418	0.916	0.420	0.103	-5.5
	0.100	3339	0.370	0.265	0.088	119	0.550	0.146	1.278	0.942	0.321	0.105	+7.6
31-1	0	445	0.455	0.207	0.094	150	0.528	0.109	1.0	1.0	0.150	0.094	-
31-2	0.284	3339	0.370	0.371	0.137	354	0.465	0.173	1.791	0.881	0.447	0.110	-19.7
	0.150	3339	0.370	0.294	0.109	241	0.490	0.144	1.418	0.928	0.252	0.112	+2.8
	0.100	3339	0.370	0.265	0.098	208	0.500	0.133	1.278	0.947	0.193	0.109	+11.2
31-1	0	445	0.455	0.207	0.094	311	0.477	0.099	1.0	1.0	0.050	0.094	-
31-2	0.284	3339	0.370	0.361	0.137	734	0.425	0.154	1.791	0.891	0.149	0.136	-0.7
	0.150	3339	0.370	0.294	0.109	500	0.445	0.131	1.418	0.933	0.084	0.120	+10.1
	0.100	3339	0.370	0.265	0.098	431	0.455	0.121	1.278	0.954	0.064	0.113	+15.3

assumption of "characteristics," lengths, and also upon assumed relationships between residence time, furnace volume, and flow rates (Eq. (26)). In the subsequent sets of calculations in Table 11, the effect of different values of $(\beta t)_b$ is demonstrated. As (βt) decreases, the experimentally determined value for $\sqrt{c^{T2}}$ remains constant, but the predicted value, based upon Eq. (40), increases. At a value for $(\beta t)_b$ of 0.05, the experimental and predicted values for unmixedness factor agree (based upon x calculated from carbon dioxide ratios).

The other variables in Eq. (40), such as the assumed values for CO/CO_2 and (A_f/A_j) , are much less important and have not been considered in detail.

Table 11 shows the relative agreement between $\sqrt{c^{T2}}$ determined from the stack gas analysis (based upon Δy calculated from Eq. (49)) and the predicted value for $\sqrt{c^{T2}}$, based upon Eq. (40). Note that for a given value of x , there is a value for $(\beta t)_b$ which will produce close agreement between the two $\sqrt{c^{T2}}$ values. Likewise, for a given value for $(\beta t)_b$, there exists a value for x which will also produce a close agreement between the experimental and theoretically predicted values for $\sqrt{c^{T2}}$. In Figure 27, the value of $(1-x)$ is shown as a function of the value of $(\beta t)_b$ required to reconcile the test data for Tests 31-1 and 31-2 with the values calculated from Eq. (40). The range of mutually consistent values for $(1-x)$ and $(\beta t)_b$ falls within the accuracy of the experimental data. Because of the uncertainty associated with both x and βt , one cannot conclude that the validity of the Corrsin theory is "proved," but only that the data are consistent with the theory to the extent that the results of the tests could be reasonably explained. Similar analyses could be made on each of the other series of tests with some notable exceptions. In Test 32 the addition of jet air caused an increase in CO_2 in Zone 1. A possible explanation for this is that the jet air was introduced at a much higher static pressure than was the air which leaks into the furnace. The introduction of jet air could change the static pressure pattern within the furnace to the extent that one mole of jet air displaces more than one mole of leakage air. The result was a decrease in the total amount of air entering the furnace and a corresponding increase in the CO_2 content. In this particular case then, the emissions are not dependent so much upon the rate of decay as upon the basic change in combustion conditions within the furnace itself.

In Test 33-2 the furnace temperature was intentionally held constant. To achieve this, it was necessary to cut down on the underfire air to the furnace. In this case also, the poor test results resulted from significant changes in the combustion conditions within the furnace rather than from the effects of turbulent mixing.

Finally, reductions in combustible emissions in Zones 2 and 3, of the breech were rarely seen. We propose two reasons for this phenomenon:

- The gases in Zones 2 and 3 emanate from portions of the furnace not covered by overfire mixing jets, and

Table 11. RELATIVE AGREEMENT BETWEEN $\sqrt{c^{(2)}}_{\text{expt}}$
AND $\sqrt{c^{(2)}}_{\text{calc.}}$

$\bar{x}$	$(\beta t)_b$		
	0.25	0.15	0.05
0.284	-35.8	-19.7	-0.7
0.150	-5.5	+2.8	+10.1
0.100	+7.6	+11.2	+15.3

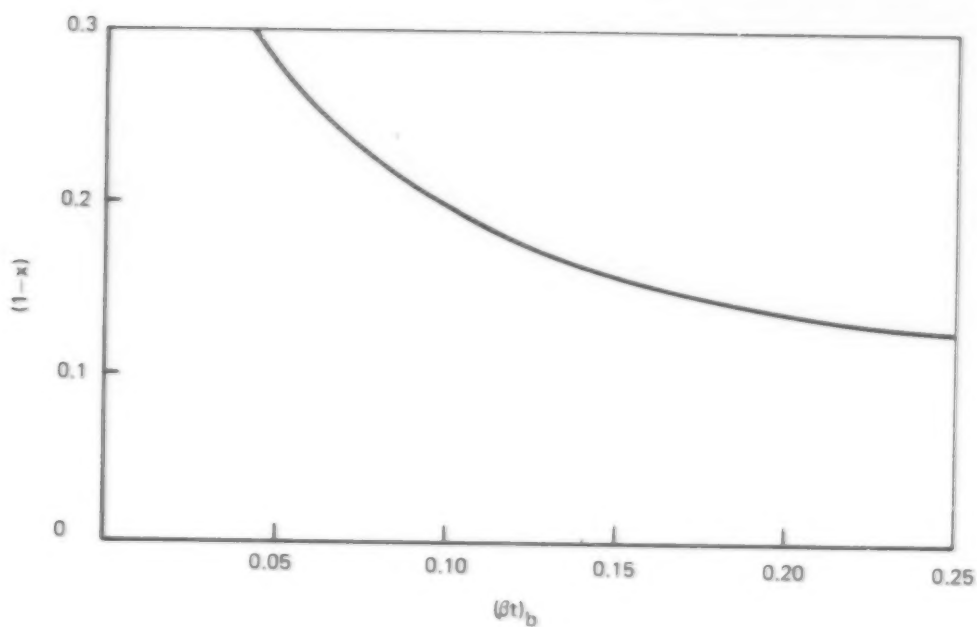


Figure 27 $(1-x)$ vs. $(\beta t)_b$

- The temperatures in these zones are lower than in Zone 1 to the extent that a significant portion of the concentration distribution could extend into the quench zone. Unfortunately, the installation of overfire mixing jets in these colder portions of the furnace would not be effective since the positive effect of the increase in mixing intensity would be offset by the negative effects of extending an even greater portion of the concentration distribution into the quench zone.

RESULTS OF PARTICULATE TESTS

As a complementary portion of the test program, pollutant emissions were measured in the incinerator stack for particulates, HCl, SO_x, and NO_x. The particulates were analyzed for trace metal content and polynuclear aromatic hydrocarbons.

The emissions were measured using the standard particulate sampling techniques (also called Method 5) as described in the Federal Register of December 23, 1971. Measurements were made during the time of the first five BASE tests. The gaseous species (HCl, SO_x, NO_x) were measured in the duct leading to the base of the stack using standard impinger collection techniques. Particulates were collected at a point in the stack 60 ft. (18m) above the foundation of the stack. Pertinent stack data at the sampling point follow:

stack O.D.	20 ft (6.1m)
stack I.D.	12 ft (3.6m)
diameters upstream	10
diameters downstream	3.4

Because of the location of the existing sampling ports, measurements were made by traversing the stack on one diameter from both sides; a total of 40 traverse points were used. The location of these points in each radius is given in Table 12.

Results of the gaseous pollutant measurements are given in Table 13. The data are all normalized to 12% CO₂. The values are expected to vary with refuse composition as observed, but there is a very wide range in the HCl values from 220 ppm.

The basic stack particulate and sampling data are given in Table 14. The particulate data are summarized in Table 15 on the basis of 12% CO₂. The amount of particulate below the 5- to 10-μ cyclone cutoff and collected on the filter is 83%, 80%, 84%, and 78% in test samples 2-5, respectively.

Table 12. SAMPLING POINTS PER RADIUS

<u>Point Number</u>	<u>Distance from Inner Wall</u>
1	5 ft 0-5/8 in
2	4' 4-3/8"
3	3' 10-1/8"
4	3' 5-7/8"
5	3' 1-7/8"
6	2' 10-3/8"
7	2' 7"
8	2' 4"
9	2' 1-1/4"
10	1' 10"
11	1' 8"
12	1' 5-3/8"
13	1' 4-1/8"
14	1' 0-7/8"
15	10-5/8"
16	8-5/8"
17	6-5/8"
18	4-1/2"
19	2-3/4"
20	7/8"

Table 13. GASEOUS POLLUTANT EMISSIONS
(at 12% CO₂)

<u>Species</u>	<u>Concentration (ppm) for Test</u>				
	<u>1</u>	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>
NO _x	130	85	220	125	170
SO _x	1,170	530	240	970	38
HCl	28	220	81	8	2

Table 14. STACK PARTICULATE DATA

	Test			
	2	3	4	5
Volume Samples (m ³)	4.63	2.94	2.12	3.10
Stack Velocity (ft/sec)	25	14	15	14
Stack Temperature (°F)	389	450	380	470
Stack Static Pressure (in H ₂ O)	0.02	0.02	0.02	0.01
Moisture (%)	21	33	28	30
CO ₂ (%)	2.7	4.1	3.1	5.0
Particulate Weights (g)				
Probe + Cyclone	0.127	0.178	0.044	0.160
Filter	0.634	0.172	0.225	0.580
Aqueous Residue ^a	0.161	0.108	0.070	0.045
Organic Extract ^a	0.112	0.103	0.018	0.032

a. of impingers

Table 15. PARTICULATE ANALYSES
(at 12% CO₂)

Portion	Particulate Weight (g) for Test			
	2	3	4	5
Probe + Cyclone	0.572	0.516	0.171	0.384
Filter	2.850	2.060	0.878	1.390
Aqueous Residue	0.725	0.303	0.273	0.108
Organic Extract	0.505	0.300	0.072	0.077
Total Probe + Cyclone + Filter	3.422	2.576	1.049	1.774
Particulate Concentration — based on Total of Probe + Cyclone + Filter				
g/m ³ (uncorrected)	0.16	0.30	0.13	0.24
g/m ³ at 12% CO ₂	0.71	0.87	0.51	0.58

Analysis for trace metals in each of the filter and probe plus cyclone fractions was carried out by spark emission spectroscopy. These results are given in Table 16 for the filter portion and Table 17 for the probe plus cyclone portion. There is a fair degree of similarity between the samples of each type, but some large differences in the relative compositions of the filter and probe plus cyclone catches.

Table 16. TRACE METALS ANALYSIS OF FILTER CATCH
(values in ppm of particulate)

Element	Base			
	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>
Al	22,000	39,000	—	52,000
Ag	30	50	40	30
B	—	—	—	—
Ba	560	3,500	2,300	2,300
Ca	5,600	7,800	9,100	2,900
Cd	250	400	200	1,700
Cu	140	200	200	170
Cr	30	80	20	10
Fe	80	200	100	100
Ga	10	20	10	10
K	42,000	58,000	11,000	20,000
Mg	800	1,500	2,300	3,500
Mn	300	400	500	900
Mo	20	60	—	10
Ni	3	10	30	20
Pb	17,000	23,000	23,000	20,000
Sn	1,700	2,700	5,500	4,000
Si	560	1,200	3,600	2,900
Sr	30	80	30	20
Ti	100	200	—	—
V	5	10	—	—
Zn	25,000	35,000	82,000	69,000
Na	42,000	58,000	11,000	8,600

Table 17. TRACE METALS ANALYSIS OF PROBE AND CYCLONE CATCH
(values in ppm of particulate)

Element	<u>Base</u>			
	<u>2</u>	<u>3</u>	<u>4</u>	<u>5</u>
Al	100,000	100,000	100,000	30,000
Ag	30	100	100	100
B	30	30	30	
Ba	1,000	1,000	1,000	1,000
Ca	100,000	100,000	100,000	30,000
Cd	30	100	100	100
Cu	300	1,000	3,000	300
Cr	300	300	300	300
Fe	100,000	100,000	100,000	100,000
K	3,000	10,000	10,000	3,000
Mg	10,000	10,000	10,000	3,000
Mn	1,000	1,000	1,000	300
Na	3,000	3,000	3,000	3,000
Ni	100	100	100	100
P	3,000	3,000	3,000	1,000
Pb	1,000	30,000	10,000	10,000
Sb	—	1,000	300	300
Sn	100	1,000	1,000	300
Si	30,000	30,000	30,000	10,000
Sr	30	100	100	30
Ti	3,000	10,000	10,000	10,000
Zn	10,000	30,000	30,000	30,000

SECTION VIII

DESIGN GUIDELINES

Up to now the report has been focussed on the description of the combustion process using various mathematical models to delineate the fluid flow and patterns within the furnace, the effect of different types of mixing, and the importance of air distribution and temperature control. All of these major topics can be focussed directly on the design guidelines for incinerators aimed at reducing combustible emissions and improving the combustion efficiency. The important parameters are furnace temperature, furnace design, and the air distribution and flow rate. We will concentrate first on these parameters and then specifically on the criteria for the design of an overfire jet mixing system to take advantage of the several benefits such a system offers.

TEMPERATURE

The plot of emission factor as a function of temperature, which was given in Figure 1, identified two phenomena which could cause significant levels of combustible emissions:

- quenching, and
- oxygen deficiency resulting from a mixing limitation.

Quenching became prominent at temperatures below 1200°F, while the limitations in high temperature mixing were observed at temperatures above 1800°F. There was apparently no difference in the combustible emission levels in the range from 1400°F to 1800°F. As a basic operating guideline, the breeching should therefore be maintained within this temperature range at all times.

These temperatures correspond to a range in CO₂ concentration of approximately 6 to 10% and a range in excess air between 100 and 200%. The latter figure can be placed in perspective by considering the recommended percent excess air requirements for efficient combustion of other fuels as shown in Table 22.

Table 18. EXCESS AIR REQUIREMENTS FOR EFFICIENT COMBUSTION

Fuel	Recommended Excess Air (%)
Natural Gas	5- 10
Fuel Oil	10- 20
Coal	30- 60
Refuse	100-200

FURNACE DESIGN

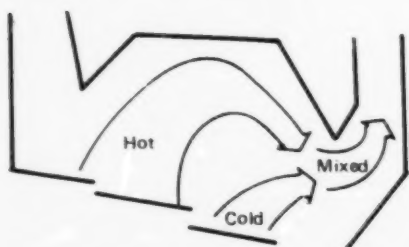
The strategy for combustible emissions is dependent upon maintaining control of the various operating parameters, relying on design parameters such as furnace configuration and relative dimensions. The designer can considerably enhance the control and mixing

required to reduce emissions. The furnace should be air-tight to avoid in-leakage that can quench the combustion or alternatively avoid smoke escaping from the furnace chamber. In sealing the furnace, however, care should be exercised to ensure that the grates are still readily accessible for maintenance.

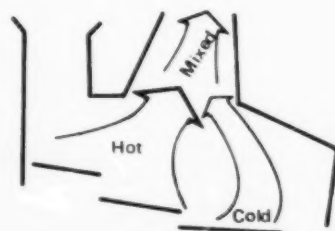
Configuration

One of the important functions of the furnace configuration with respect to eliminating combustible emissions is to ensure that the gases are uniformly mixed throughout the furnace to minimize stratification such as that observed at Newton. In our tests, the data clearly showed that the hot gases from the burning zones formed a stable layer in the upper portion of the breech, while the colder gases from the burnout zone stayed at the bottom of the breech. This was evidenced by the existence of temperature profiles within the breech and the existence of high H_2O concentrations in the colder gases caused by water evaporation in the ash quench hopper. The stratification is indicative of the existence of large temperature gradients in the furnace and suggests that zones within the furnace exist where quenching does occur. The test data clearly showed that there was a quench zone over the burnout grate and indeed high levels of combustible emissions were measured. To avoid this, the configuration of the furnace should be such as to induce front to back mixing, forcing the hotter gases to pass through the colder gases or vice versa. Several designers have suggested furnace configurations specifically aimed toward these ends.

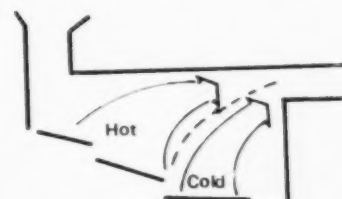
Two configuration attempting to achieve this type of mixing are shown below in comparison with the Newton configuration.



Hot gases forced through cold gases



Cold gases forced through hot gases



Hot and cold gases remain stratified

Height

The height of a furnace should be as close to the refuse surface as possible without effecting flame impingement on the roof. There are three reasons for this constraint:

1. The buoyancy of the flame causes the gases to accelerate at a rate which is dependent upon the height. The lower roof will effect a lower gas velocity, permitting the overfire air stream to influence the overall mixing patterns more strongly. This would result in better mixing control rather than

allowing gases from the bed to dominate the flow patterns in the furnace completely.

2. The lower velocity makes it easier to achieve a full jet penetration of the furnace without introducing so much air that the flame would be quenched. This is important in adding flexibility to the jet design.
3. The lower gas velocities result in smaller negative pressures at the base of the flame and therefore minimize leakage.

Furthermore, we believe that the height of the roof above the refuse bed should remain approximately constant throughout the length of the furnace. On many furnaces the roof is at a constant height, but each grate is successively lower than the one preceding it so that there is a great difference in the height above the grate at the front and the back of the furnace. This results in the possibility of establishing pressure gradients within the furnace. We observed in-leakage over the third grate at times when there was out-leakage over the first grate which confirms this concern.

Length

The length of the furnace is one of the primary parameters used in setting the refuse residence time. To achieve a good burnout, we believe that residence times on the order of 1 hour are required. Shorter residence times could be used if burning rates higher than 60 lb/hr/ft² were attained. In many configurations the length of the furnace also is the predominant factor in determining the gas residence time. The required value for the gaseous residence time is discussed below in conjunction with the jet design; times on the order of 1 second are not unreasonable in large incinerators. Smaller units having much higher mixing intensities have demonstrated effective residence times as low as 0.5 second.

Width

The width of the furnace is of primary importance in specifying jet penetration discussed below, but it is more often set to attain a desired feed rate for a given furnace length and burning rate.

UNDERFIRE AIR DISTRIBUTION AND FLOW RATE

The primary air flow in the furnace is the underfire air. The capacity of the underfire air system should be at least 150% of stoichiometric air requirements with the flexibility of being able to deliver 100% of stoichiometric requirements under any given section of the grate, especially in the sections of the grate where most of the burning occurs.

Each section of the furnace should have two important design features. First, the air plenum should have the capability to subdivide and individually control the air rates and,

second, the grates should be designed to allow an even distribution of air over the entire surface of the grate. The latter requires that the wind box be sealed so that high undergrate pressures can be attained, and that the grate have a high pressure drop with respect to refuse. In Newton, the grates have a relatively low pressure drop and the non-uniform distribution of underfire air was clearly seen. In particular, we observed the channelling of air around rather than through the bed.

The theoretical air requirements for various fuels, including refuse, are given in Table 23. As a rule of thumb, 10 scf of air are required per 1000 Btu's of heat liberated. The total air requirement is calculated from the equation:

$$Q = \frac{10 \text{ scf}}{1000 \text{ Btu}} (\text{HHV}) (\% \text{ stoic}) (\text{TPD}) \frac{2000 \text{ lb/ton}}{1440 \text{ min/day}} \quad (50)$$

where

$$\begin{aligned} Q &= \text{fan output} \\ \text{HHV} &= \text{refuse heating value (approximately 4500 Btu/lb).} \end{aligned}$$

For example, a 250-TPD furnace requires 23,400 scfm at 150% of stoichiometry (50% excess air). The air handling capacity of the grate is set by the burning rate rather than the feed rate:

$$q = \frac{10 \text{ scf}}{1000 \text{ Btu}} (\text{HHV}) (\% \text{ stoic}) \left[\frac{\text{lb/hr-ft}^2}{60} \right] \quad (51)$$

Table 19. THEORETICAL AIR REQUIREMENTS OF REFUSE AND OTHER FUELS

Fuel	lb Air/lb Fuel	lb Air/MM Btu	scf Air/M Btu
Refuse	3.22	724	9.77
Wood	3.29	703	9.49
Peat	2.33	722	9.75
Lignite	5.27	746	10.07
Sub-Bituminous B	7.58	740	10.00
Bituminous (High Volatile)	9.08	742	10.01
Bituminous (Volatile)	10.99	760	10.26
Anthracite	9.23	831	11.22
Fuel Oil	13.69	750	10.13
Methane	17.26	722	9.75

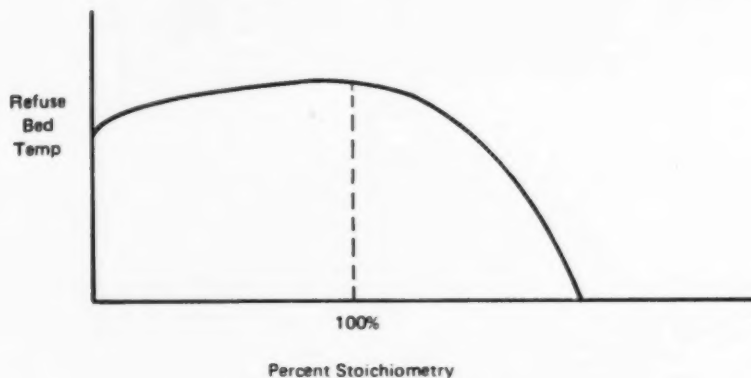
For a burning rate of 60 lb/hr-ft^2 , the underfire rate is 68 scfm/ft^2 or about 8000 scfm under a 8- by 15-foot grate. Burning rates much higher than this are possible and the designers must ensure that the grate is not limited by gas handling capacity. A safety factor of 5 applied to the burning rate, i.e., assuming a rate of 3000 lb/hr-ft^2 , is not unreasonable in the absence of actual performance data for a particular design. The impact of this design strategy is felt in these areas: burning rates, temperature control, and mixing.

Burning Rates

Our test data showed that the burning rate tended to remain constant at approximately 60 lb/hr-ft^2 , the reason being that the bed requires a minimum amount of oxygen to burn and naturally will draw in air from overfire to make up oxygen not supplied with the underfire air. As the underfire air rate increased, the overfire air entrainment decreased. We concluded that the 60 lb/hr-ft^2 was a minimum burning rate resulting from the fire's natural tendency to draw in oxygen to sustain combustion. The relatively constant burning rate exhibited for less than stoichiometric air flows confirms the existing design rule for setting burning rates at 60 lb/hr-ft^2 , but it does not suggest that this burning rate is fixed regardless of operating conditions. With positive control and adequate underfire air fan capacity, reliance on the overfire air is minimized and the burning rate is no longer limited by naturally available overfire oxygen. As a result higher burning rates are attainable. Not enough data are available to predict limiting burning rates, but values as high as 300 lb/hr-ft^2 have been reported.

Temperature Control

Underfire flow rates greater than the stoichiometric requirement also enhance the capability to control furnace temperature, particularly within the bed and at the grate. Consider the schematic diagram shown below representing the refuse bed temperature as a function of stoichiometric air flow rate. At air flows less than the stoichiometric requirements all of the oxygen will be consumed, resulting in the generation of combustion products at the adiabatic flame temperature. The gases leave the bed at approximately the



flame temperature — less any cooling of the gases — as they pass through the refuse bed. The occurrence of overfire entrainment compounds the problem by tending to stabilize the heat release rate, regardless of the underfire air setting.

The small slope of the curve above indicates that low underfire air rates provide very little refuse bed temperature control. On the other hand, for flow rates greater than the stoichiometric requirement, the additional air tends to dilute the combustion products so that an increase in the air flow results in a fairly rapid decrease in the bed temperature. This type of temperature control would be of considerable importance to the incinerator operator, but the concept is not currently in practice except in the newest units which provide adequate underfire air. Care must be exercised, however, since a large amount of excess air promotes quenching and particulate entrainment, particularly over the burnout section of the grate. The latter concern is minimized because most incinerators will soon have high efficiency air pollution control devices, and also because oxygen provided by overfire air currents tends to entrain particulate.

Mixing

Control of the underfire flow rate permits much better temperature control over the bed and effectively eliminates large variations in temperature. As a result quenching zones are eliminated and stratification is minimized.

OVERFIRE AIR JETS

The advantages of using overfire air jets are temperature control, increased mixing, and possibly slag reduction. During the course of this work it was very apparent that the sidewall jets were much more effective in controlling temperatures in the 1400°F to 1800°F range than were the existing roof jets, the reason being that jets, when properly designed, attain a full penetration of the furnace gases and result in a much more uniform mixing than do the roof jets. In addition, the jets can be effective in inducing a uniform side-to-side mixing, which is easily observed as well as increasing the mixing intensity. However, because of the furnace dimensions we believe that front-to-back mixing will be more readily achieved by proper furnace design. During the tests we were not able to demonstrate a significant increase in mixing intensity; however, from these tests we have derived criteria for evaluating the types of furnace conditions under which overfire air jets could be effective in inducing the fine-scale turbulence required to improve the combustible emissions burnout.

Additional Uses for Jets

Finally, others have suggested that jets can also be effective for reducing slag formation on incinerator walls. We were unable to confirm this observation. Certainly the early tempering of the gases with side wall jets rather than roof jets favors reduced slag formation, but the increased momentum of the particles induced by the high jet velocities can easily

increase the collection efficiency of slag on impact with the wall. Hence, care should be exercised in placing the jets to avoid inducing impingement of entrained particulate on furnace walls.

We did observe an increased rate of slag buildup directly over the jet opening and recommend the jets be surrounded by a slag-resistant refractory, such as silicon carbide.

Design Criteria

The two parameters of importance in designing overfire air jets are the jet penetration and the position of the jets. The method of Ivanov for determining jet penetration was verified as a result of the testing program. With respect to the jet configuration or spacing, several variations were tried including both opposed jets, interlaced jets, and jets acting only on one side of the furnace; all were equally effective. An additional point, not considered in earlier studies is that the jets should be located as close as possible to the burning surface for the following reasons:

- a. During the tests one could clearly observe a reduction in the flame height as a result of the overfire air jets. In furnaces designed with relatively low roof heights, as we have recommended, this aspect is very important for limiting the amount of flame impingement on the roof, thereby avoiding potential refractory damage.
- b. The temperature control and uniform mixing are more easily achieved when the gas velocity off the refuse bed is relatively low. This permits a greater control of the macroscale mixing and also results in the attainment of a full penetration with a minimum of overfire air so that quenching can be avoided.

The procedure for designing jets is outlined as follows:

- a. Determine overfire air requirement based upon temperature control –

Example:

Furnace size – 250 TPD

Stoichiometric air – 15,600 scfm

Underfire air (approximately 150%) – 23,400 scfm

Overfire air (approximately 100%) – 15,600 scfm

- b. Set jet size and number –

In practice, the jet diameter is limited in size only by the tendency of the jet to slag over. A secondary effect on jet power output, as discussed below, has

a significant effect only on small-diameter jets, but the design is usually constrained by the slagging problem. A 3-inch jet diameter is about the smallest practicable. Jet spacing should be between 3 and 5 jet diameters.

Example:

Furnace length – 40 feet

Jet diameter – 4 inches

Spacing – 16 inches

Number = 40 feet/1.33 feet/jet = 30 jets

c. Set jet penetration and configuration –

The various possible configurations – jets on one side, jets opposed on both sides, jet interlaced on both sides – appeared to be equally effective. Ivanov has suggested that jets on only one side have a slight advantage. The selection of a configuration will depend upon considerations such as slag reduction.

Example:

Configuration – one side

Penetration (furnace width) – 8 feet

d. Determine jet velocity

Since the total overfire air and the number of jets are specified above, the jet velocity can be calculated. Using Eq. (6), the required crossflow velocity can be calculated as a function of penetration:

$$u_j = Q_T / N \left(\frac{\pi d_j^2}{4} \right) = \frac{15,600 \text{ scfm}}{\pi (30) (60) \left(\frac{1/3 \text{ ft}^2}{4} \right)} \approx 100 \text{ ft/sec.} \quad (52)$$

e. Calculate required crossflow velocity (Eq. (4)):

$$u_c = 1.6 \frac{u_j d_j}{L_p} \sqrt{\frac{r_f}{T_a}}^* = 1.6 \frac{(100 \text{ ft/sec})(1/3 \text{ ft})}{8 \text{ ft}} \sqrt{\frac{2260^\circ \text{R}}{540^\circ \text{R}}} = 13.7 \text{ ft/sec} \quad (53)$$

- f. Estimate jet height above refuse (Eq. [9]):
 Temperature ratio* - $2260^{\circ}\text{R}/540^{\circ}\text{R} = 4.2$

$$u_c = Q/A_g = \frac{23,400}{(40' \times 8')60} = 1.2 \text{ ft/sec.}$$

$$\Delta z = (u_1^2 - u_0^2)/2g \left[\frac{T_H}{T_C} - 1 \right] \quad (54)$$

$$= [(13.7)^2 - (1.2)^2 \text{ ft}^2/\text{sec}^2] / 2.32 \text{ ft}^2/\text{sec}^2 (3.2) = 0.9 \text{ ft}$$

The outline above illustrates only one method for arriving at a consistent jet design. Based upon an initial air flow criterion, the order of the calculations could be changed to base the design procedure on a mixing criterion, considering the primary jet parameter to be jet penetration rather than total overfire air rate. An example in which this design approach would be more useful would be designing the mixing system for controlled-air incinerators where the underfire air is severely limited so that the introduction of large amounts of overfire air does not result in quenching.

Mixing Intensity Criteria

One of the methods for reducing combustible emissions is through an increase in the intensity of turbulence in the overfire region. The power required to attain a given level of mixing depends upon the relative densities of the fluids being mixed, but more importantly upon the size of the furnace. As a general rule, the required mixing power scales with the fifth power of the characteristic dimension of the system. Equation (24) gave the relationship for β in terms of air flow rates.

The total jet flow rate is determined as a product of the individual jet flow rate (Eq. (44)) times the grate area. In most cases the grate area is simply the product $L_p \cdot L_f$. Substituting in these values the total flow rate $Q_{T,j}$ is determined from Equation (55):

$$Q_{T,j} = \frac{\pi}{6.4 \cdot S} (L_f L_p) - u_c \sqrt{T_o/T} \quad (55)$$

The above relationship is dependent upon such things as the spacing parameter and the furnace dimensions, but is not dependent upon the jet diameter or jet velocity. Hence, in a properly designed jet system the total air flow rate introduced via overfire air jets depends

*Note: In a sealed furnace the temperature ratio becomes $2260^{\circ}\text{R}/1460^{\circ}\text{R}$ or 1.55 so that:

$$u_c = 8 \text{ ft/sec.}$$

$$\Delta z = [(8^2 - 1.2^2) \text{ ft}^2/\text{sec}^2] / 2 (32 \text{ ft/sec}^2) 0.55 = 1.8 \text{ ft}$$

upon the furnace and combustion parameters, but not on the jet design itself; i.e., the total flow rate of the jets is fixed for a given application. The power output of the mixing jets, however, is a function of the jet design, specifically being inversely proportional to the cube root of the jet diameter. For example, if a jet system is designed using 1-inch jets rather than 4-inch jets, the power output is increased by a factor of approximately 1.6. In most cases, the jet design will be constrained more by the occurrence of slagging preventing the use of small-diameter jets than by the microscale power input considerations. For this reason and others the microscale considerations do not play an important role in the jet design itself. However, the microscale relations can be of considerable importance in determining which furnace designs could benefit from the use of overfire air jets for microscale mixing and which furnaces would benefit from overfire jets only on the macroscale; i.e., for increased temperature control, slag reduction, and reduced flame heights. To be effective on the microscale the factor βt should have a value of approximately 1/2. The procedure for determining the potential effectiveness of overfire jets is as follows:

a. Determination of $\beta_0 t$

The initial value for the mixing intensity is dependent upon the furnace configuration and dimensions based upon an application of Eq. (36). The example considered below illustrates the use of these equations.

Example:

Consider the furnace jet system in the example given under *design criterion* above:

$$\beta_0 t = 0.15 \gamma^{2/3} \quad (56)$$

$$\beta_0 t = 0.15 \left\{ \frac{(2700 \text{ ft}^3/\text{sec}) (1 \text{ sec})}{(320 \text{ ft}^2) 8 \text{ ft}} \right\}^{2/3} \cong 0.19 \quad (57)$$

b. Determine mixing intensity with jets:

$$\beta t = \beta t_0 \left[1 + \left(\frac{T_f}{T_j} \right) \left(\frac{A_f}{A_j} \right)^2 \left(\frac{Q_j}{Q_f} \right)^3 \right]^{1/3} \quad (58)$$

$$\beta t = (0.19) \left[1 + 4.1 \left(\frac{320 \text{ ft}^2}{2.6 \text{ ft}^2} \right)^2 \left(\frac{270 \text{ ft}^3/\text{sec}}{2700 \text{ ft}^3/\text{sec}} \right)^3 \right]^{1/3}$$

$$\beta t \cong 0.76.$$

In this case, the incinerator would benefit from an installation of jets.

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APPENDIX A

TEST MEASUREMENT AVERAGES

Within this appendix, the averages of all the test measurements made during the testing program have been tabulated. The analytical data are listed in Table A-1. A summary of the various operating conditions under which each test was run is given in Table A-2 for the Series 20 tests, and in Table A-3 for the Series 30 tests. Finally a summary of the material and energy balance is given in Table A-4.

In addition, we have included actual test data from the initial four-week series of tests during which time the analytical procedures were being verified. These data are summarized in Table A-5. During these tests we noted severe probe leakage problems and difficulties in measuring velocity data. Therefore, the absolute quantities listed in the table are often incorrect, as can be noted by comparing the reported value of CO_2 (corrected for moisture) with the data given in Figure 26. However, the relative values should still be accurate and can be used as additional data for verifying such things as the existence of the water gas shift equilibria. Abnormally low velocities (less than $\sim 35\text{ft/sec}$) are a result of velocity probe plugging. Data in Figure 28 can be used to obtain converted readings. Finally, the measured H_2O was usually lower than the actual moisture content during the first weeks of testing before the new analytical technique for H_2O was used. Data in Figure 25 can be used to obtain corrected readings for the value. We have not relied on these data for confirming the various models presented, but we have included them in this report because much of the data is valid; considering the tremendous lack of actual test data on a municipal incinerator, even inconsistent data (when recognized as such) convey more information than has been available previously.

Table A-1. ANALYTICAL DATA

Run	DATA BREECH ZONE 1					BREECH ZONE 2					BREECH ZONE 3	
	% (Wet)			PPM (Wet)								
	CO ₂	O ₂	H ₂ O	CH ₄	CO	CO ₂	O ₂	H ₂ O	CH ₄	CO	CO ₂	O ₂
BL - 22/1	10.77	6.07	14.50	5.13	0.85	5.38	13.54	7.10	9.29	455.21	0.71	18.82
JET - 22/1	8.73	8.95	12.20	3.29	0.21	5.56	11.83	6.80	12.11	270.28	2.21	18.03
BL - 22/2	10.43	7.08	15.20	11.87	0.84	7.00	13.00	7.00	12.00	142.50	1.54	18.72
BL - 23/1	9.76	8.36	13.80	12.28	103.44	5.46	14.36	7.30	10.66	308.22	0.25	19.65
JET - 23/1	7.25	11.02	10.04	1.19	52.64	4.74	14.69	5.50	9.92	359.10	0.12	20.06
BL - 23/2	8.90	7.85	16.20	1.46	245.11	3.84	14.32	7.00	10.23	1,185.75	0.60	19.94
JET - 23/2	5.70	11.33	9.50	0.90	119.91	4.75	13.14	5.80	16.95	1,071.52	0.77	19.19
BL - 24/1	6.61	10.44	14.40	2.56	79.18	6.11	11.83	2.40	12.36	272.51	0.26	18.55
JET - 24/1	9.39	7.73	14.00	3.44	47.30	6.09	11.46	9.50	15.15	9.05	1.83	16.35
JET - 24/2	8.71	9.14	10.80	0.89	40.14	5.79	12.19	7.30	9.73	11.58	1.06	17.38
BL - 24/2	10.03	6.53	16.20	1.25	341.48	5.77	11.93	8.40	15.57	11.45	1.44	16.94
BL - 25/1	11.80	5.40	12.90	22.21	641.27	8.15	11.11	11.40	14.39	338.89	2.50	17.88
JET - 25/1	8.35	7.11	14.30	26.99	679.17	6.74	13.34	9.50	8.37	259.43	3.28	17.76
JET - 25/2	7.74	11.80	7.80	1.38	64.54	5.81	11.19	5.90	5.17	275.24	1.79	19.33
BL - 25/2	8.45	10.78	11.90	1.05	64.31	3.99	17.14	4.90	7.84	461.23	1.17	19.95
BL - 31/1	11.11	6.16	13.20	40.10	145.56	7.33	10.49	9.50	28.96	162.90	2.65	16.38
JET - 31/1	7.96	9.66	10.50	2.50	30.43	5.09	13.15	7.35	4.16	187.61	1.25	18.10
JET - 31/2	8.22	9.28	11.60	0.88	17.68	5.85	12.63	7.10	25.54	278.70	1.61	17.99
BL - 32/1	9.27	6.82	14.65	9.96	187.77	7.76	10.56	9.70	6.32	31.60	2.26	14.90
JET - 32/1	9.60	6.98	12.70	2.18	165.87	7.91	8.89	11.10	5.33	120.01	3.30	13.78
JET - 32/2	8.05	8.05	13.40	4.15	157.61	6.59	9.98	10.25	9.87	229.49	2.31	15.33
BL - 33/1	10.54	6.25	14.30	21.00	241.67	6.56	12.02	7.50	10.82	300.62	2.06	17.00
JET - 33/1	8.82	8.38	12.65	8.03	63.32	6.76	11.26	9.85	8.11	234.39	3.09	16.21
JET - 33/2	9.08	6.45	15.10	267.43	382.05	6.93	11.04	8.70	219.12	314.98	2.98	16.12
BL - 34/1	9.30	7.02	12.20	46.09	210.72	6.88	10.92	8.20*	5.69	183.60	1.63	16.09
JET - 34/1	8.63	8.37	12.75	6.63	47.98	5.85	11.53	8.45	11.62	244.89	1.41	15.97
JET - 34/2	8.00	9.82	9.00	3.18	40.95	5.04	12.62	6.50	9.81	250.11	1.90	16.40
JET - 34/3	9.20	7.17	8.00	3.68	41.40	6.50	11.33	7.10	12.07	222.96	1.34	16.86

* Approximate-based on estimates due to plugging of velocity measurement probe during a run.

Table A-1. ANALYTICAL DATA

BREECH ZONE 3					TEMPERATURE (°F)				BREECH FLOW			
CO ₂	O ₂	H ₂ O	CH ₄	CO	Zone I	Zone II	Zone III	Avg.	Velocity (ft/sec)			Total lb. moles/hr
									I	II	III	
0.71	18.82	3.10	4.84	77.76	1,721	1,446	1,029	1,366	51.80	48.28	42.78	7,643
2.21	18.03	3.80	6.73	50.50	1,508	1,387	1,004	1,272	50.41	46.58	44.77	8,007
1.54	18.72	3.40	14.94	119.52	1,700	1,458	1,100	1,411	54.17*	47.66	38.87	7,376
0.25	19.65	4.10	2.87	9.59	1,715	1,446	1,046	1,372	53.02*	51.74	44.23	7,935
0.12	20.06	2.10	3.18	9.79	1,437	1,125	737	1,058	44.51*	44.54	37.08	8,102
0.60	19.94	2.70	10.37	608.12	1,717	1,350	871	1,265	52.12*	49.05	40.81	8,026
0.77	19.19	2.80	20.16	510.30	1,421	1,158	856	1,123	48.19*	49.01	39.81	8,439
0.26	18.55	4.50	6.44	23.87	1,647	1,385	892	1,271	55.74	47.54	42.26	8,210
1.83	16.35	5.10	23.96	249.11	1,694	1,470	1,152	1,424	56.50	50.84	45.59	7,929
1.06	17.38	3.40	21.01	217.35	1,551	1,341	949	1,258	51.42	44.52	40.23	7,742
1.44	16.94	3.70	35.39	300.93	1,730	1,426	1,071	1,382	53.93	47.15	44.08	7,705
2.50	17.88	5.50	30.71	571.72	1,900	1,696	1,283	1,598	54.94*	54.86	48.24	7,496
3.28	17.76	5.40	42.57	339.37	1,796	1,587	1,225	1,505	51.38	52.53	48.57	7,587
1.79	19.33	3.20	24.20	307.34	1,400	1,246	875	1,143	47.50*	44.45*	42.22	8,186
1.17	19.95	2.20	22.49	501.22	1,523	1,131	742	1,097	49.31	44.41	36.30	8,142
2.65	16.38	5.30	38.44	181.82	1,750	1,543	1,193	1,477	53.26	48.91	43.93*	7,369
1.25	18.10	3.20	5.42	20.32	1,507	1,261	836	1,179	52.28	47.91	38.24	8,231
1.61	17.99	4.80	4.76	19.04	1,525	1,308	1,025	1,267	44.20	42.85	38.25	7,086
2.26	14.90	5.65	32.07	264.18	1,825	1,707	1,381	1,621	54.55*	51.22	48.64	7,258
3.30	13.78	8.10	34.46	22.97	1,683	1,667	1,408	1,576	51.33	52.42	48.76	7,324
2.31	15.33	7.60	62.64	200.60	1,698	1,609	1,344	1,538	51.10	50.28	46.95	7,260
2.06	17.00	6.05	42.93	420.42	1,777	1,545	1,251	1,503	48.02	46.25	42.79	6,824
3.09	16.21	6.25	76.87	517.96	1,666	1,485	1,240	1,446	45.77	43.73	42.78	6,783
2.98	16.12	6.80	950.64	829.48	1,753	1,555	1,306	1,512	45.45*	43.00	46.69*	6,723
1.63	16.09	3.60	17.83	337.40	1,793	1,478	1,150	1,458	53.93*	47.43	41.61	7,280
1.41	15.97	5.50	28.06	304.76	1,554	1,417	1,106	1,354	49.85*	42.43	37.19	6,984
1.90	16.40	4.60	73.93	453.15	1,497	1,367	1,023	1,277	48.85*	44.86	41.10	7,584
1.34	16.86	4.20	21.07	411.94	1,597	1,481	1,115	1,398	49.93	48.65	35.52	7,023

Table A-2. SUMMARY

		Purpose	Procedure	CFM	Operating Parameters			Wet/Dry	Refuse Feed Rate (TPH)	Grate Speed	
					Under Fire Air I (%)	Under Fire Air II (%)	Under Fire Air III (%)				
BL	22/1	Exploratory runs	No sidewall or roof jets	8400	11	30	59	0	Dry	9.8	45
Jet	22/1	Exploratory runs	Sidewall jets on — alternative opposed; 50% Penetration	7400	10	32	58	4900 SJ	Dry	9.8	45
BL	22/2	Exploratory runs	No sidewall or roof jets — same as BL 22/1	7400	10	32	58	0	Dry	9.8	45
BL	23/1	Exploratory runs	No roof or sidewall jets	9800	18	31	51	0	Dry	8.74	40
Jet	23/1		Alternate, opposed sidewall jets on-penetration ~ 60%	9600	17	31	52	4300 SJ	Dry	8.74	40
BL	23/2		Same as BL 23/1 except smaller UF air	9300	17	32	51	0	Dry	8.74	40
Jet	23/2		Same jet condition as Jet 23/1 except more overfire and less UF air	9100	16	35	49	5700 SJ	Dry	8.74	40
BL	24/1	Exploratory runs	No sidewall or roof jets	8400	14	37	49	0	Dry	9.78	45
Jet	24/1		Alternate opposed jets, lower UF air from BL 24/1	7000	11	39	50	2800 SJ	Dry	9.78	45
Jet	24/2		Same jet settings as Jet 24/1	7250	10	41	49	2800 SJ	Dry	9.78	45
BL	24/2		Same as Jet 24/2 except no jets on	7250	10	41	49	0	Dry	9.78	45
BL	25/1	Exploratory runs	No jets sidewall or roof	8450	9	43	48	0	Dry	9.78	45
Jet	25/1		Alternate opposed sidewall jets or penetration ~ 25%	7450	10	43	47	2200 RJ	Dry	9.78	45
Jet	25/2		Stone feed rate and raise to the jet and UF in Penetration 40%	8250	9	41	50	3600 SJ	Dry	8.74	40
BL	25/2		Same as Jet 25/2 except no jets	8250	9	41	50	0	Dry	8.74	40

SJ = sidewall jets
 RJ = roof jets
 NJ = no data
 TPH = tons per hour
 FPH = feet per hour
 CFM = cubic feet per minute

Table A-2. SUMMARY OF VARIOUS OPERATING CONDITIONS - SERIES 20 TESTS

Refuse Dry	Feed Rate (TPH)	Grate Speed (FPH)			Res. Time (hr)	Average Breach Temp. (°F)	Analysis of Operation		Relative Variation	Average Stack Temperature (°F)	Burnout Efficiency (10 ⁶ Btu/hr)	Run Times (hr)
		I	II	III			High Furnace Temperature (°F)	Low Furnace Temperature (°F)				
	9.8	45	45	45	.933	1365	1900	1640	small-med	370	88.4	1345→1445
	9.8	45	45	45	.933	1272	1670	1480	med	370	86.0	1545→1645
	9.8	45	45	45	.933	1407	1940	1550	large	370	88.0	1700→1730
	8.74	40	45	50	.928	1369	2000	1600	med-large	370	92.0	1015→1120
	8.74	40	45	50	.928	1067	1540	1260	large	370	72.6	1200→1300
	8.74	40	45	50	.928	1265	1960	1500	large	360	86.9	1330→1430
	8.74	40	45	50	.928	1122	1470	1270	small	350	77.3	1500→1610
	9.78	45	56	55	.835	1271	1760	1470	large	360	90.4	1140→1240
	9.78	45	50	55	.835	1424	1720	1520	med	360	98.4	1330→1430
	9.78	45	50	55	.835	1258	1620	1400	med-large	360	81.8	1500→1600
	9.78	45	50	55	.835	1382	1860	1530	large	370	92.7	1700→1800
	9.78	45	50	55	.835	1596	ND	ND	ND	400	104.3	1050→1150
	9.78	45	50	55	.835	1504	ND	ND	ND	400	99.3	1220→1325
	8.74	40	45	50	.928	1143	ND	ND	ND	375	76.8	1455→1555
	8.74	40	45	50	.928	1097	ND	ND	ND	380	73.4	1650→1755

Table A-3. SUMMARY OF VARIOUS OPERATING CONDIT

		Operating Conditions											
		Purpose	Procedure	CFM	Under Fire Air			Jets CFM	Refuse Wet/Dry	Feed Rate (TPH)	Grate Speed (FPH)		
					I (%)	II (%)	III (%)				I	II	III
BL	31/1	Determine effect of blocks of jets over drying grate & 2nd grate	No sidewall or roof jets run normal conditions	6400	11	39	50	0	Dry	8.74	40	45	45
Jet	31/1	Demonstrate effect of sidewall jets over the drying grate and drop zone to the second grate	Jets 1-5 on both sides penetration ~50%	6400	11	39	50	5535 SJ	Dry	8.74	40	45	45
Jet	31/2	Demonstrate effect of sidewall jets over the 2nd grate	Jets 6-10 or, both sides penetration ~50%	6400	11	39	50	4750 SJ	Dry	8.74	40	45	45
BL	32/1	Determine effect of placement of jets on roof as opposed to wall and effect of reducing feed, U Fair, and SJ air by same proposition Demonstrate effect of roof jets	Open roof jets, half open	6800	23	31	46	1200 RJ	Dry	9.78	45	45	40
Jet	32/1	Demonstrate effect of sidewall jets	Use sidewall jets – alternate opposed – both sides	7200	22	29	49	4195 SJ	Dry	9.78	45	45	45
Jet	32/2	Demonstrate effect of sidewall jets with lower flow, lower UFair, and lower feed rate	Reduce feed grate speed by 10%, reduce SJ air by 10%, reduce UF air by 10%	6750	22	31	47	3790 SJ	Dry	8.74	40	40	45
		Determine effect of one sided vs opposed jets											
BL	33/1	Determine the effects of jet placement – all on one wall or placed on both walls	Baseline measurement Use no side or roof jets	7000	21.43	33.57	45.00	0	Dry	8.74	40	50	45
Jet	33/1	Demonstrate effect of all jets on one wall	Use all jets on North wall of incinerator	6900	23.19	31.16	45.65	4260	Dry	8.74	40	50	45
Jet	33/2	Demonstrate effect of jet on both walls	Use alternate opposed jets configuration	6900	23.19	31.16	45.65	4260	Dry	9.78	45	50	45
BL	34/1	Determine the effect of jet penetration of the operation of the furnace	Baseline measurement no side jets, no roof jets,	7850	20	32	48	0	Dry	8.74	40	50	45
Jet	34/1	Demonstrate effect of partial penetration	alternate opposed jets on all jets set to 25% penetration	7100	20	30	50	2466	Dry	8.74	40	50	50
Jet	34/2	Demonstrate effect of total non-overlapping penetration	alternate opposed jets on all all jets set at 50% penetration	5450	22	31	47	4487	Dry	8.74	40	50	50
Jet	34/3	Demonstrate effect of maximum possible penetration	alternate opposed jets on jets over 2nd grate on maximum, jets over drying grate & fall area on about 3/4 (50% penetration)	7100	20	30	50	4880	Dry	8.74	40	50	50

ND-No Data

OF VARIOUS OPERATING CONDITIONS - SERIES 30 TESTS

Refuse	Feed Rate (TPH)	Grate Speed (FPH)			Res. Time (hr)	Average Breach Temp. (°F)	Analysis of Operation		Relative Variation	Average Stack Temperature (°F)	Burnout Efficiency (10 ⁶ Btu/hr)	Run Times (hr)
		I	II	III			High Furnace Temperature (°F)	Low Furnace Temperature (°F)				
	8.74	40	45	45	.96	1460	1850	1400	large	320	98.2	1105→1215
	8.74	40	45	45	.96	1179	1620	1290	large	330	80.2	1425→1535
	8.74	40	45	45	.96	1267	1560	1385	med	320	75.1	1645→1715
	9.78	45	45	40	.98	1619	1890	1660	med	340	100.5	1040→1150
	9.78	45	45	45	.93	1576	1750	1670	small	340	100.9	1345→1415
	8.74	40	40	45	1.00	1538	1840	1560	med-large	360	96.8	1545→1655
	8.74	40	50	45	.93	1503	1840	1640	small	300	87.9	1030→1140
	8.74	40	50	45	.93	1446	1690	1550	small	300	84.5	1250→1350
	9.78	45	50	45	.90	1522	1910	1590	large	320	93.1	1630→1700
	8.74	40	50	45	0.93	1455	1890	1660	med	320	88.6	1100→1150
	8.74	40	50	50	0.89	1354	1630	1470	small	320	81.2	1400→1500
	8.74	40	50	50	0.89	1277	1610	1400	med	310	79.7	1530→1600
	8.74	40	50	50	0.89	1398	ND	ND	ND	ND	80.8	1700→1720

Table A-4. SUMMARY OF MATERIAL AND ENERGY

	CARBON BALANCE			HYDROGEN BALANCE				OXYGEN BALANCE			
	lb moles/hr		% Deviation	lb moles/hr		% Deviation		lb moles/hr			
	C _{In}	C _{Out}		H _{In}	H _{In}	H _{Out}		O _{In}	O _{In}	O _{In}	
	As Refuse	As CO ₂	C _{In}	As Refuse	As H ₂ O	As H ₂ O	H _{In}	As Refuse	As H ₂ O	As O ₂	As H ₂ O
BL - 22/1	425	400	4.4	625	525	1,200	- 3.6	250	260	3,000	
JET - 22/1	425	425	- 0.8	650	550	1,150	3.5	260	270	3,200	
BL - 22/2	425	450	- 6.8	700	575	1,250	2.8	280	290	2,900	
BL - 23/1	375	375	- 1.5	600	500	1,250	- 16.8	240	250	3,100	
JET - 23/1	375	300	21.1	475	375	900	- 5.8	180	190	3,200	
BL - 23/2	375	325	12.8	500	425	1,300	- 36.6	200	210	3,300	
JET - 23/2	375	300	19.2	300	475	950	- 12.5	190	200	3,400	
BL - 24/1	425	325	21.9	525	425	1,450	- 53.0	200	210	3,200	
JET - 24/1	425	450	- 6.6	700	575	1,500	- 15.5	280	290	3,100	
JET - 24/2	425	375	8.3	600	500	1,050	2.6	240	250	3,100	
BL - 24/2	425	425	- 0.5	650	550	1,400	- 14.9	260	270	3,000	
BL - 25/1	425	525	- 28.0	850	700	1,450	6.5	330	350	2,800	1
JET - 25/1	425	450	- 6.9	700	575	1,400	- 9.7	280	290	2,900	
JET - 25/2	375	400	- 5.1	625	500	900	22.1	240	260	3,300	
BL - 25/2	375	350	7.0	550	450	950	3.0	220	230	3,200	
BL - 31/1	375	500	- 34.0	800	650	1,350	7.2	310	330	2,900	1
JET - 31/1	375	375	- 1.3	600	500	1,100	- 2.8	240	250	3,300	
JET - 31/2	375	350	5.0	550	450	1,100	- 5.7	220	230	2,800	
BL - 32/1	425	450	- 7.8	700	575	1,400	- 9.4	280	290	2,800	
JET - 32/1	425	500	- 19.0	775	650	1,550	- 7.6	310	320	2,800	1
JET - 32/2	375	400	- 6.9	625	525	1,500	- 30.0	250	260	2,800	
BL - 33/1	375	425	- 11.7	650	550	1,250	- 2.9	260	270	2,600	
JET - 33/1	375	400	- 9.6	650	525	1,250	- 7.7	250	270	2,600	
JET - 33/2	425	400	3.5	625	525	1,200	- 4.7	250	260	2,600	
BL - 34/1	375	425	- 13.0	650	550	1,150	6.0	260	270	2,900	
JET - 34/1	375	375	1.2	600	475	1,250	- 17.0	230	240	2,700	
JET - 34/2	375	375	1.8	575	475	1,000	5.3	230	240	3,000	
JET - 34/3	375	400	- 7.6	625	525	900	21.8	250	260	2,800	

A-4. SUMMARY OF MATERIAL AND ENERGY BALANCE

Deviation H _{Out} H _{In}	OXYGEN BALANCE							ENERGY BALANCE		
	lb moles/hr						% Deviation O _{In} O _{Out} O _{In}	oF/% CO ₂	[mean] % Deviation From Mean [mean] - [mean]	Heat Release In Btu/hr
	O _{In}	O _{In}	O _{In}	O _{Out}	O _{Out}	O _{Out}				
	As Refuse	As H ₂ O	As O ₂	As CO ₂	As O ₂	As H ₂ O				
3.6	250	260	3,000	800	2,000	600	2.1	260	0.9	8.9
3.5	260	270	3,200	850	2,100	575	3.2	240	- 6.6	86
2.8	280	290	2,900	900	1,800	625	2.3	240	- 9.2	88
16.8	240	250	3,100	750	2,300	625	- 4.6	290	11.6	92
5.8	180	190	3,200	600	2,600	450	.8	290	10.7	73
6.6	200	210	3,300	650	2,400	650	2.0	310	20.6	87
12.5	190	200	3,400	600	2,500	475	4.9	320	21.3	77
33.0	200	210	3,200	650	2,300	425	1.8	320	23.5	90
15.5	280	290	3,100	900	1,900	475	3.0	260	- 2.1	98
2.6	240	250	3,100	750	2,000	525	6.4	360	- 1.9	81
14.9	260	270	3,000	800	1,900	700	3.7	350	- 2.8	92
6.5	330	350	2,800	1,050	1,800	725	1.5	230	- 13.2	104
9.7	280	290	2,900	900	2,000	700	- 4.4	260	- 1.3	99
22.1	240	260	3,300	800	2,400	450	3.5	240	- 8.4	76
3.0	220	230	3,200	700	2,700	475	5.9	260	- 0.7	73
7.2	310	330	2,900	1,000	1,800	675	-	220	- 14.3	98
2.8	240	250	3,300	750	2,300	550	4.0	260	- 1.0	80
5.7	220	230	2,800	700	1,900	550	1.6	250	- 2.3	75
9.4	280	290	2,800	900	1,600	700	5.8	260	1.1	101
7.6	310	320	2,800	1,000	1,500	775	7.0	230	- 10.4	101
30.0	250	260	2,800	800	1,700	750	4.0	280	8.0	97
2.9	260	270	2,600	850	1,700	625	2.0	250	- 5.0	88
7.7	250	270	2,600	800	1,700	625	.6	240	- 7.5	84
4.7	250	260	2,600	800	1,600	600	2.2	250	- 4.3	93
6.0	260	270	2,900	850	1,700	575	10.2	260	- 2.0	88
17.0	230	240	2,700	750	1,700	625	5.6	260	- 1.0	81
5.3	230	240	3,000	750	2,000	500	8.0	270	1.9	79
21.8	250	260	2,800	800	1,700	450	13.3	250	- 5.7	80

Table A-5. COMPILATION OF TEST

RUN	MOISTURE	GRATE SPEED, FPM			UNDERFIRE AIR, CFM										OVERFIRE AIR, CFM	ZONE													
					BED 1			2	3	TOTAL	% TOTAL			% STOICHIOMETRIC			CO ₂		O ₂		H ₂ O		CH ₄		CO		H ₂		CO ₂
		1	2	3	1	2	3				1	2	3	TOTAL	←	→	←	→	←	→	←	→	←	→	←	→	←	→	←
1	WET	40	30	30	1300	2640	1410	5350	25	49	26	8	16	9	33	3650	11.0	8.00	15.0	1210	1880	1650	11.5	10.5					
2	WET	35	40	45	3780	7170	3830	14,780	26	48	26	25	48	26	99	6020	.80	20.0	3.0	10	65	0	.60	20.5					
2a	WET	35	40	45	3780	7170	3830	14,780	26	48	26	25	48	26	99	6020	.20	18.0	3.0	8	20	0	.20	18.0					
3	DRY	35	40	45	3780	7170	3830	14,780	26	48	26	25	48	26	99	6020	.25	19.5	3.0	3.6	10.8	0	.18	19.7					
4a	WET	35	35	35	980	4550	3230	8760	11	32	37	7	30	22	59	3140	.30	19.0	3.7	8.3	18.3	0	.20	19.0					
4b	WET	35	35	35	980	4550	3230	8760	11	32	37	7	30	22	59	3140	.40	19.2	3.0	32	59	30	.15	19.8					
5	DRY	55	55	55	915	3220	3380	7515	12	45	43	4	16	15	35	3635	.83	18.9	3.0	28	124	88	2.16	17.8					
6	DRY	55	55	55	1440	5250	4500	11,190	13	47	40	6	24	20	50	6580	.36	19.4	3.0	17	61	29	3.00	16.9					
7a	DRY	55	55	55	785	4110	3100	7995	10	51	39	3.5	18	14	35.5	2735	.45	19.4	3.0	10	24	10	6.13	14.0					
7b	WET	55	55	55	785	4110	3100	7995	10	51	39	3.5	18	14	35.5	2735	.49	19.7	3.0	12	73	47	5.80	14.6					
8	WET	35	30	30	965	4000	3200	8165	12	49	39	6	27	22	55	4335	.24	21.0	3.0	7	8	38	2.76	15.8					
9a	WET	50	50	50	3160	6120	3260	12,540	25	49	26	15	30	16	61	4850	0	20.0	3.0	8	21	16	(0)	(20.0)					
9b	WET	50	50	50	4170	7900	4220	16,290	26	48	26	20	39	21	80	4450	.14	20.0	3.0	10	15	19	(.14)	(20.0)					
10a	DRY	55	50	50	2350	5100	5630	13,080	18	39	43	10	23	25	58	11,470	.48	19.0	3.0	12	30	5	(.48)	(19.0)					
10b	DRY	55	50	50	2090	5100	10,120	17,310	12	30	58	9	23	45	77	9440	.21	20.0	3.0	1.7	20	42	(.21)	(20.0)					
10c	DRY	55	50	50	1960	4080	1970	8010	25	50	25	9	18	9	36	4200	0	20.0	3.0	15	13	21	(0)	(20.0)					
11	DRY	35	35	35	915	4535	3375	8825	10	51	39	6	30	23	59	4215	9.52	11.57	10.6	20	38	56	3.66	11.6					
12	DRY	55	55	55	785	4535	3490	8810	9	51	40	9	20	10	39	4365	12.0	7.6	16.4	740	2975	1555	7.18	12.6					
13a	DRY	45	45	45	1440	6425	6130	13,995	10	46	44	7.5	34.5	33	75	5850	7.8	11.5	9.9	195	3000	1800	8.50	10.7					
13b	DRY	35	50	55	330	1480	1410	3220	10	46	44	2	10	10	22	2000	10.7	8.1	9.9	42	236	14	5.17	14.4					
14	DRY	35	50	55	655	7445	7030	15,130	4	49	47	4	50	48	102	5850	6.9	12.4	9.4	41	324	149	5.1	15.5					

NOTE: Numbers in parenthesis were estimated from general profiles.

5. COMPILATION OF TEST DATA

ZONE II								ZONE III						RESIDENCE TIME, HRS	TEMPERATURE, °F				FLUID GAS FLOW			
CO	H ₂	CO ₂	O ₂	H ₂ O	CH ₄	CO	H ₂	CO ₂	O ₂	H ₂ O	CH ₄	CO	H ₂		I	II	III	AVG	I	II	III	TOTAL CFH x 10 ⁶
PPM →		%			PPM →			%			PPM →								FPS →			
1880	1650	11.5	10.5	22	1333	1738	14	7.00	13.0	8	338	1367	1700	1.31	1903	1850	1655	1773	21.1	17.2	34.2	5.27
65	0	.60	20.5	3	12	78	0	2.90	18.5	5	200	956	0	1.05	1343	1227	950	1180	45.0	48.7	42.0	9.67
20	0	.20	18.0	3	8.3	23	0	1.75	16.8	5	82	660	0	1.05	1374	1212	729	1136	47.0	46.9	36.4	9.27
10.8	0	.18	19.7	3	4	12	0	3.85	16.3	13	44	580	0	1.05	1785	1670	1302	1593	54.0	56.6	50.5	11.02
18.3	0	.20	19.0	3.7	8	12	30	1.8	18.0	3.7	58	577	75	1.20	1444	1228	928	1218	47.8	42.3	38.9	9.21
59	30	.15	19.8	3	32	45	41	2.73	17.4	5	493	1029	194	1.20	1297	1190	935	1130	38.5	39.4	33.7	7.95
124	88	2.16	17.8	5	42	253	217	3.45	16.4	6	62	601	363	.76	1922	1824	1511	1696	20.0	35.9	49.2	7.58
61	29	3.00	16.9	5	53	355	215	2.76	17.1	6	33	383	272	.76	1919	1844	1529	1677	15.7	74.5	51.6	6.85
24	10	6.13	14.0	9	58	495	78	4.10	15.6	9	93	723	299	.76	1579	1552	1276	1478	44.1	44.5	33.2	9.04
73	47	5.80	14.6	9	68	628	96	2.40	18.1	9	136	793	84	.76	1311	1233	986	1186	38.7	36.9	32.7	7.73
8	38	2.76	15.8	5	32	344	25	.94	15.9	3	45	342	38	1.35	1356	1044	703	1083	44.2	36.7	28.0	7.74
21	16	(0)	(20.0)	(5)	(8)	(21)	(16)	1.10	19.0	3	158	813	40	.84	1446	(1296)	985	1349	18.6	(17.9)	7.4	2.71
15	19	(.14)	(20.0)	(5)	(10)	(15)	(19)	.22	20.0	3	19	258	0	.84	1775	(1625)	1223	1641	13.4	(13.0)	3.8	2.12
30	5	(.48)	(19.0)	(5)	(12)	(30)	(5)	2.40	18.0	4	29	312	37	.82	1821	(1671)	1419	1593	24.4	(23.6)	42.7	6.58
20	42	(.21)	(20.0)	(5)	(1.7)	(20)	(42)	.21	20.0	4	0	50	0	.82	1811	(1461)	1244	1378	19.4	(18.7)	45.1	6.08
13	21	(0)	(20.0)	(5)	(15)	(13)	(21)	0	20.0	4	25	70	0	.82	1936	(1786)	1586	1709	21.0	(20.4)	51.5	6.79
38	56	3.66	11.6	6.2	24	12	38	1.65	19.4	2.5	38	228	28	1.20	1520	1218	869	1230	40.5	34.2	31.4	7.57
2975	1555	7.18	12.6	10.9	201	2750	1250	4.60	15.3	6.2	64	618	123	.76	1982	(1850)	1559	1755	26.1	19.3	39.7	6.18
3000	1800	8.50	10.7	9.9	142	2500	1800	4.50	15.9	5.1	33	500	49	.93	2030	(1920)	1629	1863	43.3	30.9	38.2	8.07
236	14	5.17	14.6	9.9	16	210	14	3.0	17.0	5.1	50	417	64	.91	1656	(1410)	1231	1379	11.4	17.0	25.0	3.86
325	149	5.1	15.5	6.2	16	270	39	2.28	18.4	4.5	41	450	56	.91	1690	(1560)	1211	1424	11.1	12.7	22.0	3.32

APPENDIX B

CALCULATION OF UNMIXEDNESS FACTOR OF REFUSE BED OFFGAS

The ratios of moles of various components with respect to moles of offgas are shown in Table B-1 below:

Table B-1. MOLE RATIOS					
	Unreacted Mixture			Reacted Mixture	
	Moles	Percent		Moles	Percent
Offgas	1	100%	Combustion* Products	1	100%
		$1 + 4.76n_{O_2} + n_{X_s}$			$1 + 3.76n_{O_2} + n_{X_s}$
Stoichiometric Air	$4.76n_{O_2}$	$\frac{(4.76n_{O_2}) \cdot 100\%}{1 + 4.76n_{O_2} + n_{X_s}}$	n_2	$3.76n_{O_2}$	$\frac{3.76n_{O_2} \cdot 100\%}{1 + 3.76n_{O_2} + n_{X_s}}$
Excess Air	n_{X_s}	$\frac{n_{X_s} \cdot 100\%}{1 + 4.76n_{O_2} + n_{X_s}}$	Excess Air	n_{X_s}	$\frac{n_{X_s} \cdot 100\%}{1 + 3.76n_{O_2} + n_{X_s}}$
Total	$1 + 4.76n_{O_2} + n_{X_s}$	100%	Total	$1 + 3.76n_{O_2} + n_{X_s}$	100%

Note that in a stoichiometric mixture n_{X_s} is zero. For one mole of offgas, n_{O_2} moles of oxygen are required and therefore $4.76 \cdot n_{O_2}$ moles of air. The mole fraction of offgas in a stoichiometric ratio is therefore $4.76 \cdot n_{O_2}$:

$$y_s = \frac{\text{moles of combustible}}{\text{total moles}} = \frac{1}{1 + 4.76 \cdot n_{O_2}} \quad (B-1)$$

The mole fraction of offgas in a typical mixture (not stoichiometric) is:

$$y = \frac{1}{1 + 4.76(n_{O_2} + n_{X_s})} \quad (B-2)$$

To relate these relations to the gas analysis, consider a blending of a stoichiometric mixture $y_s \cdot C_t$; the concentration of offgas in the excess air is zero. Since concentration, C , can be calculated from the mole fraction and the ideal gas law, then:

$$\begin{aligned} C_s &= y_s \cdot C_t = y_s \frac{P}{RT} \\ C &= y \cdot C_t = y \frac{P}{RT} \end{aligned} \quad (B-3)$$

*None of the combustion reactions involved results in a net production of mols.

For blending,

$$C_s \text{ (volume of stoichiometric mixture)} + O.D \text{ (volume of excess air)} = C \text{ (volume of total)}$$

Hence:

$$(y_s \cdot C_t)V_s = (y \cdot C_t)V_t \text{ or } y_s n_s = y n_t \quad (B-4)$$

Then, for the example in Table B-1,

$$\begin{aligned} C_t V_s &= \frac{PV_s}{RT} = n_s = 1 + 4.76 n_{O_2} \\ C_t V_t &= \frac{PV_t}{RT} = n_t = 1 + 4.76 (n_{O_2} + n_{x_s}) \end{aligned} \quad (B-5)$$

Therefore:

$$\frac{y}{y_s} = \frac{n_s}{n_t} = \frac{1 + 4.76 \cdot n_{O_2}}{1 + 4.76(n_{O_2} + n_{x_s})} \quad (B-6)$$

The initial concentration or mole fraction of offgas is not always known, and it is important to determine y/y_s in terms of gas analysis after combustion. The same approach is used as before, considering a mole of final combustion products rather than offgas. In this case:

$$n_{c_p} = n_{CO_2} + n_{H_2O} + 3.76 n_{N_2}$$

and

$$\%CO_2|_s = \frac{n_{CO_2}}{n_{c_p}} \cdot 100\% \quad (B-7)$$

$$\%CO_2 = \frac{n_{CO_2}}{n_t} \cdot 100\%$$

Hence

$$\frac{\%CO_2}{\%CO_2|_s} = \frac{n_{c_p}}{n_t} = \frac{1 + 3.76 \cdot n_{O_2}}{1 + 3.76 \cdot n_{O_2} + 4.76 n_{x_s}} \quad (B-8)$$

This can be rearranged to:

$$\begin{aligned} \%CO_2 &= \%CO_2|_s \left[\frac{(1 + 3.76 \cdot n_{O_2})}{(1 + 4.76(n_{O_2} + n_{X_s}) - n_{O_2})} \right] = \\ \%CO_2|_s &\left[\frac{(1 + 3.76 \cdot n_{O_2}) y}{1 - n_{O_2} \cdot y} \right] \end{aligned} \quad (B-9)$$

Solving for y yields:

$$y = \left[\frac{(\%CO_2 / \%CO_2|_s)}{1 + n_{O_2} (\%CO_2 / \%CO_2|_s)} \right] \left[\frac{1}{1 + 3.76 n_{O_2}} \right] \quad (B-10)$$

Then:

$$\Delta y = y_s - y = y_s \left\{ 1 - \frac{\%CO_2}{\%CO_2|_s} \left[1 + n_{O_2} \cdot \left(\frac{\frac{1}{1 + 3.76 n_{O_2}} - \frac{\%CO_2}{\%CO_2|_s}}{1 + \frac{\%CO_2}{\%CO_2|_s} \cdot n_{O_2}} \right) \right] \right\} \quad (B-11)$$

When n_{O_2} is small, the equation reduces to

$$\Delta y = y_s \left(1 - \frac{\%CO_2}{\%CO_2|_s} \right) \quad (B-12)$$

Example: For the bed offgas where $CO/CO_2 = 0.5$, $n_{O_2} = 0.058$, $y_s = 0.784$.

At a temperature of 1500°F, $\%CO_2 = 7.2\%$.

$$\Delta y = 0.783 \left\{ 1 - \frac{7.2\%}{15.1\%} \left[1 + 0.058 \left(\frac{\left(\frac{1}{1 + 3.76 (0.058)} \right) - \frac{7.2\%}{15.1\%}}{1 + \frac{7.2\%}{15.1\%} (0.058)} \right) \right] \right\} \quad (B-13)$$

$$\Delta y = 0.404$$

Using the simplified expression above, Δy is estimated to be 0.410.

The required oxygen, O_2 req, is calculated based upon CO:

$$O_2 \text{ req} = n_{O_2} \cdot (CO/CO|_s) \quad (B-14)$$

The available oxygen is equal to %O₂ from the gas analyses.

Hence:

$$\frac{O_2 \text{ req}}{O_2 \text{ available}} = \frac{n_{O_2} (CO/CO|_s)}{\%O_2} \quad (B-15)$$

For the example above, a CO concentration of 100 ppm results in a value of 8.6×10^{-4} .

The factor $\sqrt{c'^2}/\Delta y$ is obtained from Figure 3 and is approximately 0.40.

Since

$$\sqrt{c'^2} = \left(\sqrt{c'^2} / \Delta y \right) O \cdot \Delta y \quad (B-16)$$

then, the unmixedness factor for the example is:

$$0.40 \times 0.404 = 0.162$$

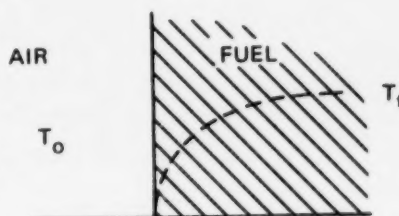
APPENDIX C

MATHEMATICAL CHARACTERIZATION OF QUENCHING

The problem of calculating the rate at which a pocket of fuel will be quenched by reduced temperature and determining the corresponding fuel concentration at the time of quenching may be modeled as described below:

The temperature of the gas is cooled by conduction of heat from the fuel at the adiabatic flame temperature to the surrounding air at ambient conditions. This represents the most rapid rate of conduction heating and other factors, such as air preheating or radiative heating, will diminish this rate as well.

Consider a portion of fuel as shown in the figure right in which the temperature is initially T_f . At time, $t=0$ the surface is brought to temperature T_0 and held constant; again this is the maximum gradient. The governing boundary value problem is:



$$\partial T / \partial t = \alpha \partial^2 T / \partial z^2 \quad (C-1)$$

$$T(0, z) = T_f$$

$$T(t, 0) = T_0, \quad T(t, \infty) = T_0$$

The solution is:

$$\left[\frac{T_f - T}{T_f - T_0} \right] = 1 - \operatorname{erf} \left(\frac{z}{\sqrt{4 \alpha t}} \right); \quad \alpha = k / \rho C p_{\text{fuel}} \quad (C-2)$$

If this is evaluated for gases at the quench temperature, then:

$$T_f = 3100^\circ\text{F}$$

$$T_0 = 100^\circ\text{F, and}$$

$$T_q = 1100^\circ\text{F (quench temperature).}$$

and, therefore:

$$\operatorname{erf} (z_q / \sqrt{4 \alpha t}) \cong 1 - (3100 - 1100 / 3100 - 100) \cong 1/3$$

$$z_q / \sqrt{4 \alpha t} = 0.305 \quad (C-3)$$

where $z_q(t)$ is the distance from the initial surface to where the reaction is quenched. The rate of propagation of the quenching front is:

$$d(z_q)/dt = 0.305 \sqrt{\alpha/t} \quad (C-4)$$

If one assumes that the fuel is initially at the adiabatic flame temperature (less a reasonable amount of heat loss) and must be cooled by dilution with ambient air, then the change in temperature and concentration will be continuous. For any combustibles to remain following this transition, the mechanism for cooling would have to precede so that the stoichiometric concentration C_s would not be attained before the fuel reached the quench temperature. The above analysis shows the rate of propagation of the quench front; the propagation of the plane where C_s is reached must be slower than $0.305 \sqrt{\alpha/t}$. Otherwise, a condition would exist in which fuel and air would exist in proper proportions and above the quench temperature; the expected result of this condition is a complete burnout of fuel.

First, consider the case of air diffusing without reaction, the analogous case to the heat condition problem. The governing equation for C_a is:

$$\begin{aligned} \partial C_a / \partial t &= D_a (\partial^2 C_a / \partial z^2) & C_a + C_f &= C_T \\ C_a(t, 0) &= C_{a0}, C_a(t, \infty) &= 0 \\ C_a(0, z) &= 0 \end{aligned} \quad (C-5)$$

The solution is:

$$(C_{a0} - C_a / C_{a0}) = \operatorname{erf}(z / \sqrt{4 D_a t}) \quad (C-6)$$

For the case where $CO/CO_2 = 0.5$, C_s was shown to be 0.78.

Then:

$$\frac{C_{a0} - C_a}{C_{a0}} = 0.78 = \operatorname{erf}\left(\frac{z_s}{\sqrt{4 D_a t}}\right) \quad (C-7)$$

in which case

$$z_s = 0.87 \sqrt{4 D_a t} \quad (C-8)$$

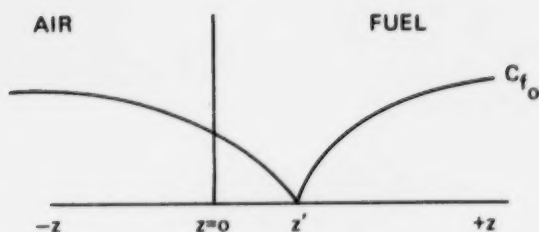
Then, when

$$\alpha = 11.2 \text{ ft}^2/\text{hr (constant)}$$

$$\dot{D}_a = 9.3 \text{ ft}^2/\text{hr (constant)}$$

$$\frac{d(zs)}{dt} = 0.79\sqrt{\alpha/t}$$

Since the stoichiometric concentration front propagates over twice as fast as the quench zone according to this mechanism, the condition of attaining C_s at $T > T_q$ is a certainty and no combustible would be expected.



Consider next the effect of a rapid irreversible reaction as shown in the figure left. Since both air and fuel cannot co-exist, there is a plane at z where the concentration of both is zero. For the constant molar case:



The governing equation for each region is:

<u>Air</u>	<u>Fuel</u>
$C_a + C_p = C_T$	$C_f + C_p = C_T$
$\partial C_a / \partial t = D_a (\partial^2 C_a / \partial z^2) \quad -\infty < z < z'$	$\partial C_f / \partial t = D_f (\partial^2 C_f / \partial z^2) \quad z' < z < \infty$
$C_a(0, z') = C_{a0}$	$C_f(0, z') = C_{f0}$
$C_a(t, z') = 0, C_a(t, -\infty) = C_{a0}$	$C_f(t, z') = 0, C_f(t, \infty) = C_{f0} \quad (C-9)$
$-D_a \frac{\partial C_a}{\partial z} \Big _{z=z'} = n \cdot D_f \frac{\partial C_f}{\partial z} \Big _{z=z'}$	

The solution is of the form

$$\begin{aligned} C_a &= a_1 + a_2 \operatorname{erf} \left(\frac{z}{\sqrt{4 D_a t}} \right) \\ C_f &= b_1 + b_2 \operatorname{erf} \left(\frac{z}{\sqrt{4 D_f t}} \right) \end{aligned} \quad (C-10)$$

Solving for the boundary values, the solution is:

$$\frac{C_{a0} - C_a}{C_{a0}} = \frac{1 + \operatorname{erf}\left(\frac{z}{\sqrt{4D_f t}}\right)}{1 + \operatorname{erf}\left(\sqrt{\eta/D_f}\right)}$$

$$\frac{C_{f0} - C_f}{C_{f0}} = \frac{1 - \operatorname{erf}\left(\frac{z}{\sqrt{4D_f t}}\right)}{1 - \operatorname{erf}\left(\sqrt{\eta/D_f}\right)} \quad (\text{C-11})$$

where

$$z' = \sqrt{4\eta t}$$

From the final boundary conditions, representing the reaction stoichiometry, the implicit relation for η is derived:

$$\frac{1 - \operatorname{erf}(\sqrt{\eta/D_f})}{1 + \operatorname{erf}(\sqrt{\eta/D_a})} = \frac{\eta \cdot C_{f0}}{C_{a0}} \sqrt{\frac{D_f}{D_a}} \exp \left\{ \sqrt{\eta/D_a} - \sqrt{\eta/D_f} \right\} \quad (\text{C-12})$$

For the case $D_f = D_a = D$, $C_{a0} = C_{f0} = 1$ (pure fluids) the relation reduces to

$$\operatorname{erf}(\sqrt{\eta/D}) = \frac{1-n}{1+n} \quad (\text{C-13})$$

Then for $\text{CO}/\text{CO}_2 = 0.5$, $n = 0.23$ mole air/mole fuel

$$z' = 0.57 \sqrt{4\alpha t}$$

$$\frac{d}{dt} (z') = 0.57 \sqrt{\alpha/t} \quad (\text{C-14})$$

Because of the small amount of air required, the value for η is still not small enough for the rate of propagation of C_s to be slower than that for T_q .

The final step in the analysis is to consider the effects of a fuel temperature lower than the theoretical adiabatic flame temperature. Such an occurrence results from two phenomena. First, the combustion reactions in the bed do not go to completion so that all of the heat of reaction is not liberated and the flame temperature is proportionately lower. The additional amount of heat is liberated as oxygen diffuses, but the net reduction in the heat liberated would occur if a pocket of combustibles were quenched before the reaction was completed.

A second mechanism for having a fuel temperature lower than the adiabatic flame temperature is a result of heat transfer from the hot combustion gases to the refuse bed itself. This is evidenced by an increase in the temperature of the refuse bed as well as an increase in the production of gasification or pyrolysis products. In the case of pyrolysis which produces large amounts of hydrocarbons, the consistency with the gasification premise for bed-burning is largely the result of the decomposition of methane to carbon and hydrogen, which is substantially complete at temperatures above a 1000°F.

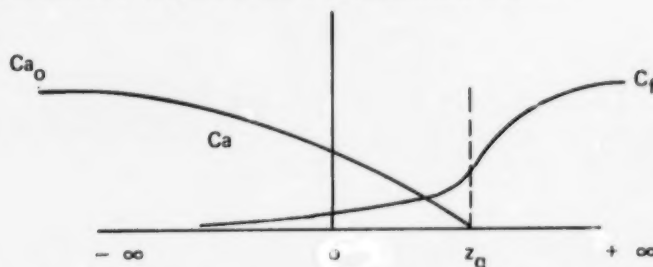
Finally, a pocket of fuel can lose heat through radiation. This mechanism tends to cool the hottest gases and heat the colder gases and will result in a net loss of heat for the high temperature fuel pockets.

If we assume that the temperature of the fuel is only 2000°F, as a result of the three mechanisms above, then the left hand side of Eq. (C-3) has the value of 0.48 rather than 0.305. Hence, the speed of propagation of the quenched plane would be greatly increased; but in this particular case, the increase would still not be sufficient to overcome the propagation of the C_s plane. However, the assumption of pyrolysis on a larger scale suggests that the ratio CO/CO_2 could be much larger than 0.5. Hence, the factor n in Eq. (C-13) would be larger and the rate of propagation for the C plane would be correspondingly slower. As an example, consider the ratio $CO/CO_2 = 1.0$. From a material balance the mols of air required practically double so that $n \cong 0.5$. Then the left hand side of Eq. (C-13) has a value of $1/3$. The solution for this value is identical to the initial example for the proportion of T_q Eq. (C-13). Under these conditions the propagation of C_s would occur at a slower rate than the propagation of the quench plane. When this condition is obtained, Eq. (C-9) would still be an adequate description of the process if the boundary conditions were changed to:

$$C_f(t, z_q) = C_q$$

$$-D_a \frac{\partial C_a}{\partial z} \Big|_{z=z_q} = n \left[D_f \frac{\partial C_f}{\partial z} \Big|_{z=z_q} + -D_f \frac{\partial C_f}{\partial z} \Big|_{z=z_q} \right] \quad (C-15)$$

The concentration profiles are shown schematically below:



where in this case z_q marks the location of the quench plane. To the right side of z_q the mixture is above the quench temperature and, therefore, the oxygen concentration must be 0. To the left side, the mixture is below the quench temperature so that the co-existence of both oxygen and fuel is possible. Under these conditions the equation for C_a is unchanged but the equation for C_f becomes:

$$\frac{C_{f0} - C_f}{C_{f0} - C_q} = \frac{1 - \operatorname{erf} \frac{z}{\sqrt{4 D_f t}}}{1 - \operatorname{erf} \left(\frac{z_q}{\sqrt{4 D_f t}} \right)} \quad z_q < z < +\infty$$

$$\frac{C_f}{C_q} = \frac{1 + \operatorname{erf} \frac{z}{\sqrt{4 D_f t}}}{1 + \operatorname{erf} \left(\frac{z_q}{\sqrt{4 D_f t}} \right)} \quad -\infty < z < z_q \quad (C-16)$$

Interestingly enough, the concentration where quenching occurs is constant as can be deduced from Eq. (C-10) for C_f by substituting $z_q = .305 \sqrt{4 \alpha t}$. If for the mixture on the right hand side of z_q the diffusion equation of the form for Eq. (C-5) is solved for fuel and products individually as well as the two combined, then it can be shown that

$$C_p/C_f = (C_p/C_f) \big|_{z=z_q} \quad (C-17)$$

The stoichiometric boundary condition in the case of quenching reflects the net consumption of fuel at the boundary z_q . This can be solved to yield:

$$\frac{1 - \operatorname{erf} \left(\frac{z_q}{\sqrt{4 D_f t}} \right)}{1 + \operatorname{erf} \left(\frac{z_q}{\sqrt{4 D_a t}} \right)} = \frac{n(C_{f0} - C_a)}{C_{a0}} \cdot \sqrt{\frac{D_f}{D_a}} \exp \left\{ \frac{z_q^2}{4 D_a t} - \frac{z_q^2}{4 D_f t} \right\} \left\{ 1 - \frac{C_q}{C_{f0} - C_q} \right\} \quad (C-18)$$

which is similar to (Eq. 12), but with the substitution of $(C_{f0} - C_q)$ for C_{f0} and with the additional term $[1 - (C_q/C_{f0} - C_q)]$ representing the flux of fuel into the quenched mixture. Equation (C-12) was used to solve for z_q . In this case, however, z_q is determined from the heat conduction problem. Substituting Eq. (C-3), for example, and assuming equal diffusivities ($D_a = D_f = D$), Eq. (C-18) can be used to calculate the quench concentration C_q . Simplifying the above,

$$C_q = \frac{1}{2} \left(C_{f0} - \frac{C_{a0}}{n} \right) + \frac{C_{a0}}{n} \left\{ \frac{\text{erf}(0.305 \sqrt{\alpha/D})}{1 + \text{erf}(0.305 \sqrt{\alpha/D})} \right\} \quad (\text{C-19})$$

which reduces to Eq. (C-13) when C_q equals zero and $C_{f0} = C_{a0} = 1.0$. For the case when $n = 0.5$, C_q has a value of only 0.034. If, however, z_q is assumed to equal $0.48 \sqrt{4 \alpha t}$, as suggested above, then for $n = 0.5$, C_q is equal to 0.21 which is substantial.

Up to now analysis has centered around showing not necessarily that quenching will always occur, but only that under the right set of circumstances it can occur. The experimental data presented in Section VII clearly show that in all three zones a certain amount of quenching is witnessed. Several additional sophistications can be added to the analysis, including the effect of differences in diffusivities of fuel and air and the effect of considering the diffusivities as functions of temperature rather than as mean average constants. The latter is of particular note because diffusivities vary with $T^{3/2}$, while the conductivity is a function of $T^{1/2}$. The basic conduction and diffusion equations reduce to:

Conduction

$$\frac{\eta}{\theta} \frac{d\theta}{d\eta} = \alpha_0 \frac{d^2 \theta}{d\eta^2}$$

$$\eta = \frac{z}{\sqrt{2t}} \quad ; \quad \theta = T^{3/2}$$

Diffusion

$$\eta \frac{dC}{d\eta} = -D_0 \frac{d}{d\eta} \left(\theta \frac{dC}{d\eta} \right)$$

Note that both the thermal diffusivity α and the molecular diffusivity D are functions of $T^{3/2}$, but because of the placement of the derivatives the differential equations are not of the same temperature functionality and the respective solutions will reflect this difference. Unfortunately, these equations are all highly non-linear and require a numerical solution rather than an analytical one. Since little could be gained from this additional degree of sophistication, the numerical analysis was not carried out.

From the above analysis, it is clear that the quenching phenomena can cause combustible emissions under the right set of circumstances so that a final refinement is required to devise a method for calculating these emissions. In reality, the transfer between the fuel phase and the air phase cannot be legitimately represented by semi-infinite medium since both of these are finite. The solutions of the basic differential equations with respect to finite boundary conditions have been worked out and will not be attempted here.

The simultaneous heat and mass transfer mechanism is not the only means by which emissions can be produced in gases having the appropriate amount of air and sufficient temperature to sustain combustion. In a recent paper by Sivashinsky and Gutfinger* the hypothesis that emissions can be caused because of the finite speed of chemical kinetics was explored. They concluded that during the course of combustion not all of the combustible gases are consumed in the reaction zone and that a portion of the combustible component passes through the flame front without reacting with oxygen.

Still others have suggested a stochastic model based upon eddy collisions and applying the same type of probability equations used in the kinetic theory of gases.

The exact mechanism or combination of mechanisms is not known and research is underway in this area. The quantification of quenching cannot proceed beyond the level given here until a better physical description is at hand.

* Sivashinsky, G. I., and Gutfinger, C., *The Extinction of Spherical Diffusion Flames*, Paper B5, Proc. 1973 International Seminar on Heat Transfer from Flames, Aug., 27-31, 1973, Trogir, Yugoslavia.

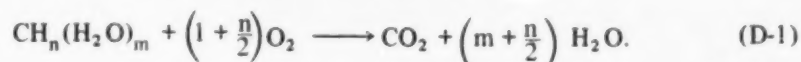
APPENDIX D

BASIC PRINCIPLES OF COMBUSTION

Within this appendix, the fundamental principles of combustion are presented for those unfamiliar with combustion chemistry.

Reaction Chemistry

The principles of combustion of refuse are no different than those of other fuels, and a great deal of study has been centered around the reactions that occur (3, 4). Most hydrocarbon fuels are parafinic and can be represented by the formula $(CH_2)_n$. Refuse is more cellulosic than parafinic and a more accurate formula would be $CH_n (H_2O)_m$. The equation for the complete combustion of refuse can then be given as:



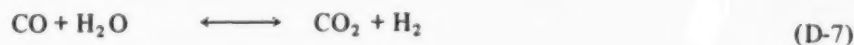
In reality the reaction does not go to completion for some fraction of their fuel and off-gas products. Partial combustion products, such as CO, H_2 , hydrocarbons, and soot, often result. The step reactions which take place either in the refuse bed or overfire region include oxidation reactions:



and reduction reactions:



By combining reactions [5] and [6], one obtains:



The last of these reactions, known as the water gas shift reaction, is of considerable importance in describing the gaseous equilibrium in bed-burning processes.

Reaction Kinetics

The reaction rate for a particular molecular species is dependent upon a combination of all the above reactions. For example, the rate of reaction for carbon monoxide is given by the equation:

$$+\frac{d(\text{CO})}{dt} = (r_4 - r_5 + 2r_7 + r_8). \quad (\text{D-8})$$

A similar equation could be derived for each of the constituents, provided that all the occurring reactions involving the constituent were identified.

The expression for the chemical reaction rate is also dependent upon the specific mechanism of reaction. In very simplified terms, the rate can be expressed by the law of mass action applied to molecular reactants. For reaction [D-3] above:

$$r_3 = k_3 (\text{CO}) (\text{O}_2)^{1/2} \quad (\text{D-9})$$

where k_3 = chemical kinetic constant, and (CO) , (O_2) = concentration of molecular species.

Unfortunately, the mechanisms for combustion reactions are rarely as simple as the elementary expressions above. Typically, the mechanisms involve such things as free-radical chain reactions or other types of activated complexes. The exact mechanisms have been worked out over years of research and will not be discussed in detail here.

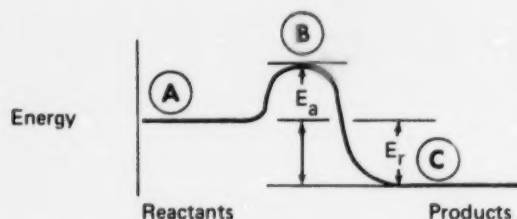
All of the oxygen-consuming reactions have extremely rapid reaction rates. This has been experimentally verified in many instances where reactions were shown to go to completion within a matter of microseconds, and certainly the rate of reaction in an internal combustion engine, or in a hydrocarbon explosion, attests to the rapidity of the chemical kinetics.

The above reactions can also go in the reverse direction, and although the same principles apply in determining the chemical kinetics for these reactions, the rate constants are different. The ratio of the rate of the forward reaction to the rate of the backward reaction determines the equilibrium of the reaction which is discussed in detail in a later section.

Activation Effects

All the kinetic expressions have their bases in statistical mechanics and represent the rate at which molecules or other reacting species will collide in such a way as to undergo a chemical change. But molecules must not only collide before a reaction takes place, they must also have sufficient energy to be able to rearrange chemical bonds to undergo the

chemical change. The large majority of molecules colliding do not have this energy, and thus the collision will not result in a chemical reaction. The process is shown schematically below:



The average molecule is assumed to have an energy level represented by point A. Upon collision, the molecule does not react. However, certain molecules, because of past history, will attain energy states higher than average. When one of these reaches the energy level of point B, then, like a roller coaster, it will seek a point of lower energy and will either return to the normal state A, or react upon collision to form products having an energy represented by point C. The height of the energy barrier E_a is the activation energy, and the difference between points A and C is the net energy of reaction, the latter being directly related to the heat of combustion.

For both the forward and reverse reactions, the dependence of the reaction rate constant was shown by Arrhenius to be exponentially dependent upon the activation energy as shown by the equation:

$$k = k_0 e^{-(E_a/RT)} \quad (D-10)$$

where E_a for the forward reaction is the difference between points A and B, while E_a for the reverse reaction is the difference between points C and B. In the case of the exothermic reaction, such as combustion, heat is liberated and the forward-reaction activation energy is less than the value for the reverse reaction, so that the rate of forward reaction is faster and the net release of energy allows the reaction to be self-sustaining once it is initiated. This suggests an additional ramification of the activation energy – the concept of ignition.

Consider a combustible mixture in which none of the molecules has sufficient energy to exceed the energy barrier for a given reaction. In this case, none of the molecules will undergo a reactive collision and combustion will not occur. As the temperature of the system is increased, the molecules increase in energy to the point where one of the molecules will gain sufficient energy to exceed the energy barrier. When this happens, a combustion reaction will occur, liberating energy, thereby increasing the temperature of the gas, and causing additional molecules to exceed the energy barrier. The chain process repeats again and again, and the temperature of the gas increases dramatically. The temperature at which this series initially begins is known as the ignition temperature.

Chemical Equilibrium

Because the chemical reactions are occurring simultaneously in both the forward and reverse directions, the reaction will ultimately never reach the state of "complete combustion." The equilibrium between the forward and the reverse reactions is determined from the ratio of the respective rate constants.

In the case of the oxidation reaction, the heat of reaction is very high and the forward reaction occurs at a much faster rate than does the reverse reaction. The reduction reaction, on the other hand, has a low heat of reaction and is, in some cases, endothermic (requires heat). As a result of the Arrhenius Equation [10], the equilibrium constant is a function of temperature according to the relation:

$$K = \frac{k_f}{k_r} = K_0 \exp ([E_{af} - E_{ar}]/RT) = \frac{P_c P_d}{P_a P_b} \quad (D-11)$$

This relation is shown for the various combustion reactions in Figure D-1. In the case of the oxidation reactions, reaction [3], for example, the equilibrium between carbon monoxide, oxygen, and carbon dioxide at 1800°F is on the order of 10^7 atm^{-1} . For a chemical equilibrium-controlled reaction in which the stack gas contains 12% CO_2 (dry) a material balance suggest approximately 8% oxygen. The concentration of carbon monoxide is calculated, below:

$$K_4 = 10^7 \text{ atm}^{-1/2} = \frac{(\% \text{ CO}_2)}{(\% \text{ CO}) (\% \text{ O}_2)^{1/2}} \frac{(100\%)^{1/2}}{(1 \text{ atm})^{1/2}} \quad (D-12)$$

$$(\% \text{ CO}) = 0.04 \text{ ppm}$$

For reaction [7] above, the equilibrium at 2000°F is equal to:

$$K_7 = \frac{(\text{CO}_2) (\text{H}_2)}{(\text{CO}) (\text{H}_2\text{O})} = 2.0 \quad (D-13)$$

Diffusion Flames

Unfortunately, the overall rate of chemical reaction cannot always be controlled by the chemical kinetics and equilibrium is often unattained. The controlling step often is the rate at which fuel and oxygen can be brought together to react. As a good example, the rate at which soot particles burn is controlled primarily by the rate at which oxygen diffuses to the surface of the soot, rather than by the rate at which carbon (soot) and oxygen combine to form carbon monoxide. Flames of this type are diffusion flames and are distinguished from premixed flames in that oxygen must diffuse into the fuel to form a combustible mixture. Examples of diffusion flames range from candles to bonfires. A distinction is made, however, in the type of diffusion flame, depending upon the type of diffusion occurring. For the small candle, the oxygen is fed in by molecular diffusion and the flame is called a laminar diffusion flame. In the case of a bonfire, the hot combustion gases have enough buoyancy to generate a considerable amount of turbulence and draw air into the base of the flame by a convective transport mechanism. These flames are turbulent diffusion flames.

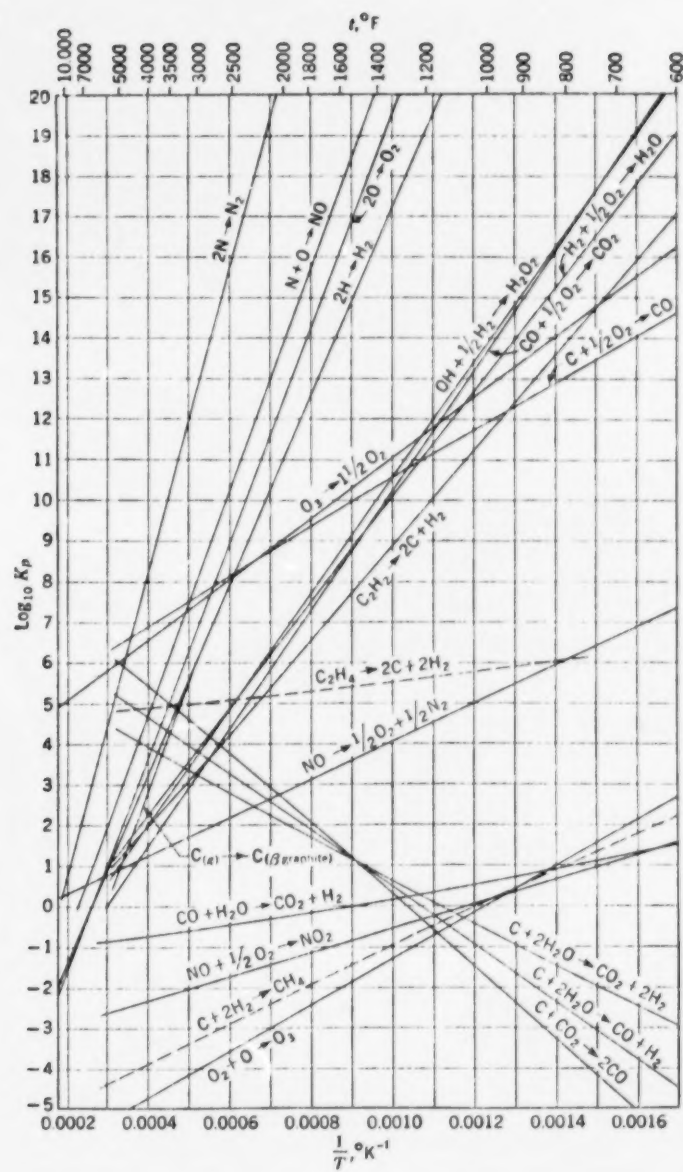


Figure D-1 Equilibrium constants of combustion reactions.

In a laminar diffusion flame the combustion is controlled by molecular diffusion described in generalized coordinates by Fick's Law:

$$j = -D \frac{\partial c}{\partial n_i} \quad (D-14)$$

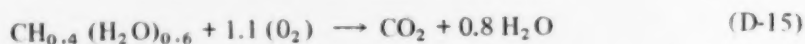
The rate of flux toward the flame boundary is directly related to the difference in concentrations at the boundary and the concentration at a distance from the boundary, i.e., the concentration gradient. The concentration profile about the flame front can be described by a mathematically continuous function analytically determined by the simultaneous solution of Fick's Law and the law of conservation of mass for a given system geometry. The solution is relatively straightforward and has been discussed at great length in the scientific journals.

In a turbulent system the solution for diffusive transport is not so easy a problem and the mathematical theory is still evolving around this issue. Unlike the laminar problems, turbulent transport takes place in bulk rather than on a molecular scale, i.e., groups of molecules tend to transfer together from one place to another via a convective transport mechanism. The process is referred to as eddy diffusion. In a turbulent regime, a whole range of eddy sizes exist and there are eddies within eddies. An adequate description of the eddy characteristics requires statistical techniques of a much greater sophistication than the analytical techniques applied to the laminar diffusion process.

The incinerator does have its own air supplies and can best be described as partially premixed turbulent diffusion flame. However, the premixing notwithstanding, the rate of gas phase combustion is controlled by turbulent diffusion and the principles describing turbulent diffusion flames will apply.

Material Balances

In several studies of the nature and composition of refuse, including those specifically related to the City of Newton's refuse, the average composition of dry ash-free refuse was approximately $\text{CH}_{0.4} (\text{H}_2\text{O})_{0.6}$ so that the combustion reaction is:



The coefficients of each reactant and product indicate the relative amounts (in mols) that are required to combine exactly during the reaction. These proportions can be converted to actual compositions or weight of burning material using the law of conservation of mass. A summary of the relative proportions of combustion products is shown in Table D-1. More often the composition of a gas sample is given in volume percent which is equivalent to the mol percent shown in the table as well. Another unit, the Orsat analysis, is usually given on a dry basis. The interrelationship between these various quantities can be summarized by the equations below applied for carbon dioxide balances. These equations, N_g , represent the

number of moles of gases in a stoichiometric mixture containing CO_2 , H_2O , and the nitrogen equivalent of oxygen required to complete combustion; i.e.:

$$n_s = n_{\text{CO}_2} + n_{\text{H}_2\text{O}} + n_{\text{N}_2,s} \quad (\text{D-16})$$

If n_{xs} represents the number of moles of oxygen present in excess of that required to complete combustion, then the total number of moles, n_t , is determined from the equation:

$$n_t = n_s + 4.76 n_{xs} \quad (\text{D-17})$$

The gas composition (is mole percent) can be calculated for CO_2 as follows:

$$(\% \text{CO}_2)_w = \frac{n_{\text{CO}_2}}{n_t} \times 100\% \quad (\text{D-18})$$

$$(\% \text{CO}_2)_d = \left(\frac{n_{\text{CO}_2}}{n_t - n_{\text{H}_2\text{O}}} \right) \times 100\% \quad (\text{D-19})$$

$$(\% \text{CO}_2)_{s,w} = \frac{n_{\text{CO}_2}}{n_s} \times 100\% \quad (\text{D-20})$$

$$(\% \text{CO}_2)_{s,d} = \frac{n_{\text{CO}_2}}{n_s - n_{\text{H}_2\text{O}}} \times 100\% \quad (\text{D-21})$$

Since 1.1 moles of oxygen are required per mole of carbon (Equation [D-15]), the percentage of excess air is calculated by the following equation:

$$(\% \text{XS}) = \frac{n_{xs}}{1.1 n_{\text{CO}_2}} \times 100\% = \frac{(\% \text{O}_2)_d}{1.1 (\% \text{CO}_2)_d} \times 100\% \quad (\text{D-22})$$

Likewise, if one assumes that the refuse initially has 0.7 inch H_2O per mole of C (35% moisture on an ash-free basis):

$$n_s = n_{\text{CO}_2} + 1.5 n_{\text{CO}_2} + 3.76 (1.1 n_{\text{CO}_2}) = 6.64 n_{\text{CO}_2} \quad (\text{D-23})$$

Finally, Equation [D-17] can be rearranged to the form:

$$\frac{n_s}{n_t} + \frac{4.76 n_{xs}}{n_t} = 1 \quad (\text{D-24})$$

Combining Eqs. (D-23) and D-24) and rearranging the fraction so that n_t and n_s are both denominators yields:

$$\frac{\frac{n_{\text{CO}_2}}{n_t}}{\frac{n_{\text{CO}_2}}{6.64 n_{\text{CO}_2}}} + \frac{\frac{n_{\text{N}_2}}{n_t}}{\frac{1}{4.76}} = 1 \quad (\text{D-25})$$

or

$$\frac{(\% \text{CO}_2)_w}{15.1\%} + \frac{(\% \text{O}_2)_d}{21.1\%} = 1 \quad (\text{D-26})$$

Table D-1. PRODUCTS OF REFUSE COMBUSTION

Fuel		Air	
Refuse	Mols	O ₂ (required)	N ₂ (3.76 X O ₂)
C	1.0	1.0	3.76
H	1.6	0.4	1.50
O	0.6	(0.3)	(1.12)
Moisture (H ₂ O)	0.7		
Total		1.1	4.14

Combustion Products				
Component	Mols	% Wet	% Dry	Weight (%)
CO ₂	1.00	15.1	19.5	23.5
H ₂ O	1.50	22.6	—	14.5
N ₂	4.14	62.3	80.5	62.0
	6.64	100.0	100.0	100.0

Energy Balances

In an adiabatic process (no external heat loss or gain to the system), a stoichiometric mixture of refuse and air reacts liberating approximately 7500 Btu/lb of dry ash-free refuse. All of the liberated heat goes to raising the temperature of the combustion products from T₀ to the adiabatic flame temperature T_a. The energy equation is:

$$(n_s \bar{C}_{p_s}) (T_a - T_0) = \sum_{i=1}^n (n C_p)_i (T_a - T_0) = 7500 \text{ Btu/lb} \quad (\text{D-27})$$

where:

n_i = mass of component i per lb of dry ash-free refuse

$n_s = \sum_{n=1}^n n_i$ (stoichiometric mixture)

C_{p_i} = heat capacity of component i

The heat capacity for the stoichiometric mixture is shown in Table D-2.

Table D-2. HEAT CAPACITIES

Component	Mols	C_p (cals/gmols - °C)	mC_p
CO ₂	15.1	13.0	196.3
H ₂ O	22.6	9.0	203.4
N ₂	62.3	7.4	461.0
Total	100.0	8.61	860.7

The final temperature of the stoichiometric mixture is referred to as the theoretical adiabatic flame temperature and is approximately 2950°F for refuse. A temperature of this intensity can damage a refractory incinerator. The primary method for reducing the temperature to a reasonable operating range is through the addition of excess air. The gas temperature is lowered because the heat liberated by combustion is used to raise the temperature of more than just the combustion products. The energy balance is as follows:

$$(n_s \bar{C}_{p_s}) (T - T_o) + (4.76 n_{x_s} \bar{C}_{p_{x_s}}) (T - T_o) = 7500 \text{ Btu/lb} \quad (\text{D-28})$$

Equating Equations D-27 and D-28 yields:

$$\left(\frac{T - T_o}{T_a - T_o} \right) = \frac{n_s \bar{C}_{p_s}}{n_s \bar{C}_{p_s} + 4.76 n_{x_s} \bar{C}_{p_{x_s}}} = 1 - \frac{4.76 n_{x_s} \bar{C}_{p_{x_s}}}{\Delta H / (T - T_o)} \quad (\text{D-29})$$

Equation [D-29] can be simplified to:

$$\left(\frac{T - T_o}{T_a - T_o} \right) = 1 - \frac{4.76 n_{x_s}}{n_s} \cdot \frac{\bar{C}_{p_{x_s}}}{\bar{C}_{p_s}} \cdot \left(\frac{T - T_o}{T_a - T_o} \right) \quad (\text{D-30})$$

Substituting Equation [D-26] and rating that n_{x_s}/n_s equals $(\%O_2)/100\%$, the equation becomes:

$$\left(\frac{T - T_o}{T_a - T_o} \right) = \frac{(\%CO_2)_w}{(\%CO_2)_{s,w}} + \frac{(\%O_2)}{(21.1\%)} \cdot \frac{(\bar{C}_{p_s} - \bar{C}_{p_{x_s}})}{\bar{C}_{p_s}} \cdot \left(\frac{T - T_o}{T_a - T_o} \right) \quad (D-31)$$

The second term on the right is a correction factor for the changes in the heat capacity in going from reactants to products and is on the order of 0.1. Since both the stoichiometric CO_2 and the adiabatic flame temperature are fixed for a given refuse, the equation predicts a linear relationship between temperature and CO_2 for a given refuse.

APPENDIX E

CHARACTERIZATION OF THE BED-BURNING PROCESS

The overall objective of developing a model of the processes occurring within the refuse bed is to determine the optimum distribution of both overfire and underfire. Previous attempts at describing the bed burning have approached the burning as a combination of several processes. The resulting models have been simplistic, providing little more than a qualitative description of the refuse burning process. The difficulty does not lie in understanding the physical processes that are occurring, but rather in mathematically combining the processes into a single coherent model.

The approach taken here is somewhat different. In the mixing study several phenomena which could occur and control the bed-burning mechanism were identified. The experimental data on burning obtained in this program allow these phenomena to be placed in perspective so that conclusions regarding design relations can be drawn.

Following a series of short tests over each of the three grates of the Newton incinerator, we selected the first three sample ports for complete bed tests. The data from these tests are given in Table E-1. These two tests and the tests made in the breech form the basis for conformation of the bed model presented here.

DESCRIPTION OF THE BURNING PROCESS

Several processes occur within the fuel bed, including drying, ignition, gasification, combustion, and char burnout. Earlier qualitative models for fuel bed combustion identified separate drying, gasification, and char-burning regions defined so that a single process was occurring in each region. However, these operations actually occur simultaneously because of the heterogeneous character of refuse.

The parameters controlling the burning bed are not well defined. Certainly one of the more important parameters is air availability which is dependent upon how much and from what region air is introduced into the furnace. Most municipal incinerators are designed to introduce the primary combustion air from beneath the refuse bed, i.e., underfire air.

Consider a fuel bed supplied exclusively with underfire air. Oxygen will be consumed as the underfire air passes through the bed so that, for underfire air rates less than that required for complete combustion, there will occur a point within the bed where the concentration of oxygen becomes zero. In a recent controlled study of incineration the existence of this point was verified.*

Table E-1. TEST RESULTS OVER BED¹

	Sample Port			
<u>Bed Test 1</u>	<u>1</u>	<u>2</u>	<u>3</u>	<u>Average</u>
CO ₂	5.93	1.31	5.8	4.35
O ₂	13.4	18.9	14.6	15.63
CO	3.2	0.26	0.13	1.20
H ₂	1.72	0.21	0.09	0.67
HC	1.16	0.01	0.03	0.40
H ₂ O ²	11.51	2.15	8.84	7.50
H ₂ O/CO ₂	1.94	1.64	1.52	1.72
CO/CO ₂	0.54	0.20	0.02	0.28
K _{wg}	3.61	2.03	2.2	3.1
Temperature	1835	1905	1928	1889
<u>Bed Test 2</u>				
CO ₂	13.3	6.4	12.0	10.6
O ₂	6.0	14.2	8.6	9.6
CO	3.35	1.00	0.26	1.54
H ₂	2.7	0.77	0.19	1.22
HC	1.94	0.34	0.39	0.89
H ₂ O	12.9	19.4	26.4	17.57
H ₂ O/CO ₂	0.97	3.03	1.70	1.66
CO/CO ₂	0.25	0.16	0.02	0.15
K _{wg}	1.20	3.94	2.33	2.09
Temperature	1787	1758	1811	1785

1. Complete sets of data are limited due to experimental difficulties directly over the bed.
2. Calculated from national balance.

In our tests, this was not the case. In both tests directly over the bed, the oxygen concentration was on the order of 10%.

The presence of air in the bed offgas could be the result of any one of the three mechanisms shown in Figure E-1.

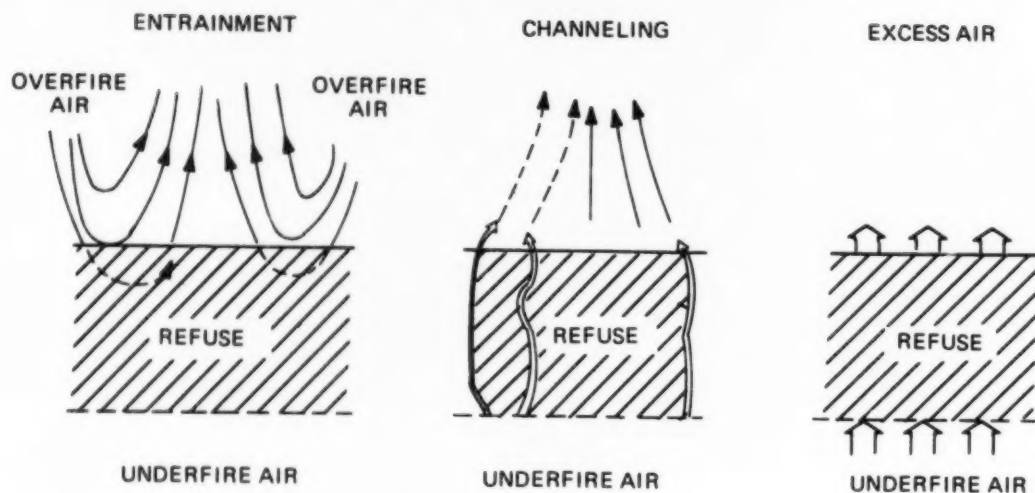


Figure E-1 Oxygen breakthrough mechanisms

Excess Air

When the refuse bed burns to a point where the kinetics for the solid combustion become very slow, then air can pass through the bed at a faster rate than that at which it can be consumed. When the amount of excess air is small compared to the amount of air that is consumed, the effect is of little consequence, since the air breaking through the bed will be well mixed with the partial combustion products, and complete burnout of the complete emissions will be eminent. However, when the fraction of oxygen breakthrough is much larger than the amount of oxygen that can be consumed by the reaction, the heat liberated by combustion will not be sufficient to raise the temperature of the gas stream and quenching will occur. The latter situation is of particular importance in the final stages of refuse burnout in which the refuse approaches a fuel-limiting condition. In the latter fuel-limiting case, the gases are uniformly mixed and the equilibrium is controlled by oxidation. In the case of both channeling and air entrainment in which non-uniform mixing exists, the water gas shift equilibrium will apply to those portions of the gas which contain combustibles but no oxygen. The non-uniformity of mixing is not unreasonable when considering the mechanism by which the oxygen breakthrough occurs.

Air Entrainment

Several recent studies have shown that air can be drawn in from the overfire region and entrained into the gases emanating from the refuse bed. In such a case, the bed-burning process must be considered as a combination of both an underfire and an overfire mechanism in which the zero-oxygen planes by each mechanism occur within the bed.

Channeling

Because of the lack of a completely random packing of the refuse, channels, or "blow holes," may develop. The "blow holes" allow the air to bypass the refuse completely. This phenomenon is a common occurrence in packed bed equipment.

Each one of these three mechanisms could control the burning. In the case of the tests at Newton, the overfire entrainment mechanism was believed to be the controlling factor. This is best illustrated by considering the magnitude of local burning rates.

BURNING RATES

The overall burning rate can be determined from the amount of oxygen consumed, from the amount of CO_2 , CO, and hydrocarbons produced, or from the amount of heat liberated. Table E-2 summarizes the overall burning rates derived from both carbon and energy balances. The averages — 54 and 67 lb/hr-ft² of grate area — based upon carbon and energy balances, respectively, are close to the value of 60 lb/hr-ft² often used as a design value. Several assumptions on bed density, ash content, moisture content, and heating value were introduced in the calculations. Errors in these assumed values may account for the difference in burning rate obtained by the carbon and energy balances.

The fraction of total air consumed which could have been introduced underfire is also given in Table E-2. Only one third of the total consumed air could have originated as underfire air. In this case, the computed burning rates are 1.7 to 2.1 times that corresponding to stoichiometric combustion of all of the underfire air.

The overfire entrainment mechanism is the only one which would result in burning rates higher than the burning rate equivalent to complete combustion of the underfire air.

The results presented in Table E-2 also indicate that the overall burning rate is not drastically affected by the underfire air rate, but remains relatively constant at about 60 lb/hr-ft² of grate area in agreement with empirical design rules used for years in designing burning surfaces. The rule holds because the total amount of air tends to remain constant. Reductions in underfire air are offset by increases in overfire air entrainment.

OFFGAS COMPOSITION

To apply the concepts of "unmixedness" and "eddy decay," it is necessary to know the initial composition of gases entering the turbulent mixing (overfire) region. From basic research in kinetics of combustion reactions, we know that oxygen and hydrogen or carbon monoxide react on the order of microseconds. For all practical purposes they cannot co-exist. However, the data given in Table E-1 show over 3% CO at times when oxygen is over 10%. One reasonable explanation for this co-existence is that the offgases are relatively unmixed so that the oxygen is not in intimate contact with the combustibles.

Table E-2. SUMMARY OF BURNING RATES AND UNDERFIRE AIR CONSUMPTION

Run No. *	Burning Rate (lb/hr-ft ²)		% of Consumed Air Introduced Underfire
	Heat Balance Total Heat 3900 x 336 **	Mass Balance CO ₂ x 23.2 (0.51) 336 †	
BL 22/1	67.9	53.3	35.5
J 22/1	65.6	56.5	31.3
BL 23/1	70.2	50.1	41.5
JET 23/1	55.3	41.5	40.6
BL 23/2	66.3	43.7	39.3
JET 23/2	59.0	40.5	38.5
BL 24/1	68.9	43.8	35.7
JET 24/1	75.1	59.8	29.6
JET 24/2	62.4	51.4	30.6
BL 24/2	70.7	56.4	30.6
BL 25/1	79.5	70.9	35.7
JET 25/1	75.8	60.0	31.5
JET 25/2	58.6	53.3	34.9
BL 25/2	56.4	46.6	34.9
BL 31/1	75.3	69.2	27.04
JET 31/1	61.5	50.8	27.04
JET 31/2	57.7	47.6	27.04
BL 32/1	77.1	59.7	28.9
JET 32/1	77.3	66.7	30.4
JET 32/2	74.2	53.5	28.5
BL 33/1	67.4	56.0	29.6
JET 33/1	64.8	54.9	29.2
JET 33/2	71.4	57.9	29.2
BL 34/1	67.6	55.6	33.2
JET 34/1	62.0	49.5	30.04
JET 34/2	60.8	49.2	23.06
JET 34/3	61.6	53.9	30.04

* Run conditions are given in Appendix A

**

†

This being the case, the characterization of the offgas composition must include the following factors:

- 1) Composition of combustible fraction,
- 2) Fraction of combustibles vs. fraction of air, and
- 3) Degree of unmixedness between two fractions.

These are discussed separately.

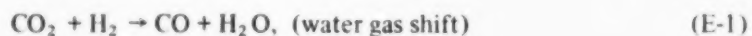
COMPOSITION OF COMBUSTIBLE FRACTION

The data gathered from earlier measurements* of bed offgas compositions suggest that once the oxygen is depleted, a burning refuse bed acts as a gasifier, producing CO, CO₂, H₂, and H₂O with minor amounts of hydrocarbons. This must occur at some place within the bed in order to have combustibles present in the offgas. The controlling equilibrium for this portion of the offgas is between these gasifier products.

Tentative bed-burning models have been developed*** on the assumption that the water gas shift reaction controls the gasification equilibrium. The experimental basis supporting the postulated equilibrium was previously limited to data taken by Kaiser† on the Oceanside (N.J.) incinerator and data taken at M.I.T. on synthetic refuse††. The data collected in this program also support this hypothesis.

The composition of the combustible fraction can be determined as a function of CO/CO₂. The calculation scheme is as follows:

For the reaction:



the equilibrium constant is defined as:

$$K_{wg} = \frac{(\text{H}_2\text{O})(\text{CO})}{(\text{CO}_2)(\text{H}_2)} = f(T) \quad (\text{E-2})$$

From the material balance:



the equilibrium concentration can be calculated as follows:

$$\text{Let: } \text{CO}/\text{CO}_2 = x, \text{ then } (\text{H}_2\text{O}) = \left(\frac{K_{wg}}{x} \right) (\text{H}_2) \quad (\text{E-4})$$

Choosing as a basis one mole of CO_2 corresponding to x moles of CO , then the total moles of carbon is $(1 + x)$. From the material balance, the total hydrogen (in the form of H_2) is determined to be $1.5 (1 + x)$, since the ratio H_2/C equals 1.5. Using the equilibrium relation above and noting the sum of H_2O and H_2 equals the total hydrogen yields $1.5(1 + x)$; then the relative proportions of H_2O and H_2 can also be calculated. Finally, the amount of N_2 present is 3.76 times the amount of oxygen consumed. The latter is equal to the total oxygen present ($\text{CO}_2 + 1/2 \text{CO} + 1/2 \text{H}_2$) minus the amount of O_2 originally contained in the initial $(1 + x)$ moles of refuse, i.e., $0.65 (1 + x)$. These proportions are tabulated below.

Relative Proportion of Combustion Products

<u>Product</u>	<u>Mols</u>
CO_2	1
CO	x
H_2O	$\frac{1.5 (1 + x) K_{wg}}{x + K_{wg}}$
H_2	$\frac{1.5 (1 + x) x}{x + K_{wg}}$
N_2	$\frac{3.76}{2} \left\{ 1 + (1 + x) \left(\frac{1.2 K_{wg} - 0.3x}{x + K_{wg}} \right) \right\}$
Total	$2.5 (1 + x) + \frac{3.76}{2} \left\{ 1 + (1 + x) \left(\frac{1.2 K_{wg} - 0.3x}{x + K_{wg}} \right) \right\}$

The compositions, determined by dividing the number of moles of each component by the total, are given in Table E-3 and are shown as a function of x in Figure E-2.

Several methods for estimating the ratio of CO/CO_2 are possible:

1. The ratio can be assumed to be equal to the corresponding ratio found for coal gasification or, more recently, in refuse pyrolysis (gasification)

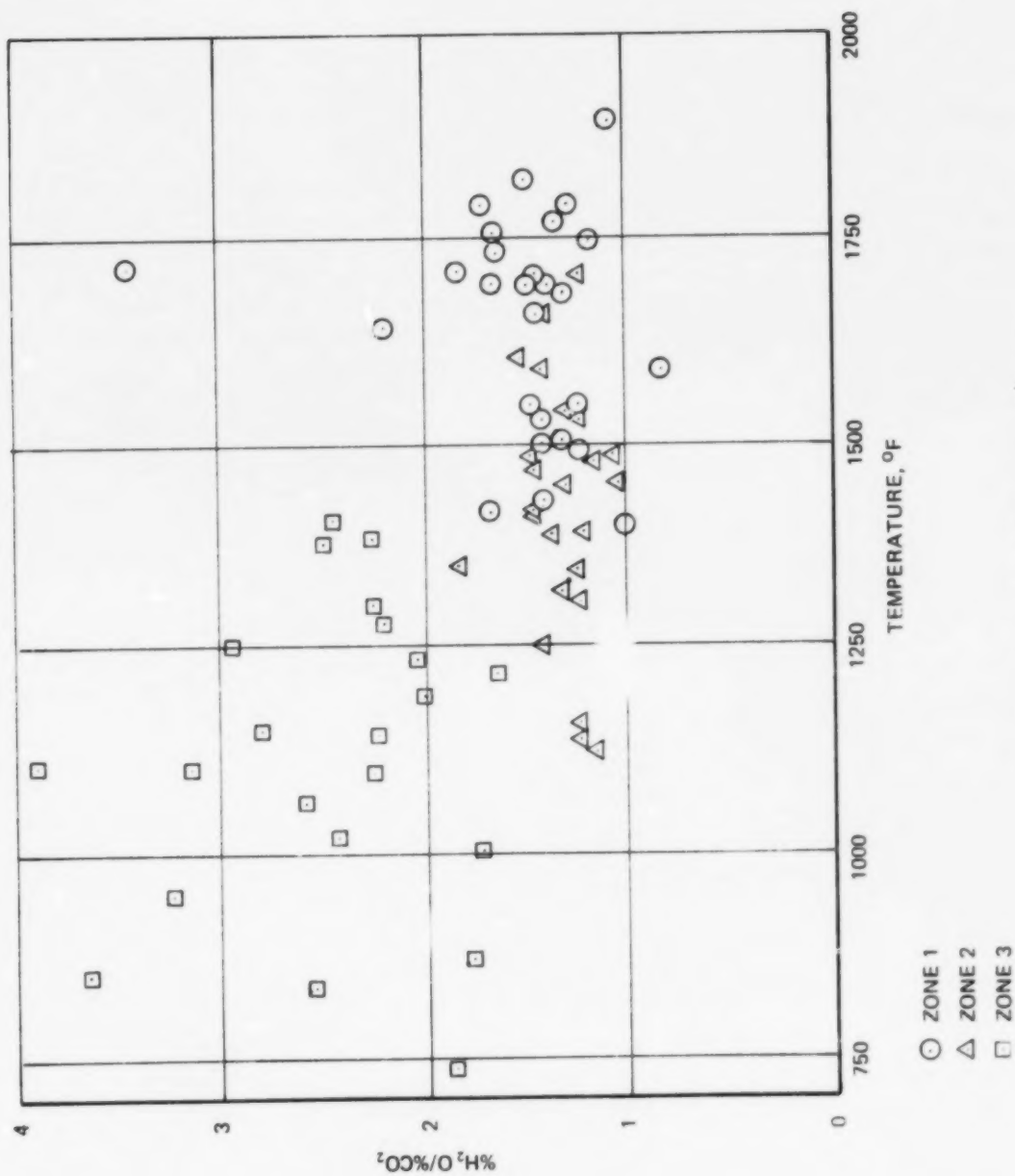


Figure E-2 Plot of $\text{H}_2\text{O}/\text{CO}_2$ versus temperature

programs. The assumption is based on the concept that the mechanism for forming CO and CO₂ in the bed of an incinerator is the same as that for gasification of other organic combustibles. This is approximately valid before combustible burnout begins in the overfire region.

2. The ratio can be determined by actual sampling over the bed. The sampling is extremely difficult and the analysis must always be suspect.

In both cases, the ratio is only a crude estimate, but the sophistication required to predict the exact carbon ratio is neither currently available nor particularly warranted.

The analysis in Section V was based upon the CO/CO₂ ratio as a characterization of the extent to which the combustion reaction has gone to completion. From the summary of the bed data in Table E-1 an empirically determined ratio ranges from 0.02 to 0.6 with the average being approximately 0.21. These values can be compared reasonably well with values obtained for coal gasification or for pyrolysis of refuse. These values are probably minimum because some overfire mixing has occurred before the gases from the bed can be sampled in the sample probe. Thus, a certain amount of reaction would be expected within the probe itself, reducing the amount of CO and correspondingly increasing the amount of CO₂ measured in each sample. Hence, the empirically determined CO/CO₂ ratio is not inconsistent with the types of initial ratios necessary to develop a condition in which quenching would dominate (Appendix C).

A second tenet of the equilibrium model is that combustion is relatively uniform throughout the bed, i.e., drying, gasification, combustion, burnout, etc, occur simultaneously. Although this is not rigorously true, the errors introduced by such an assumption are no greater than those resulting from analytical measurements.

Verification of these two points: the equilibrium of gases and the uniformity throughout the furnace is given below.

Two independent checks were made of the water-gas shift equilibrium. For the data collected in the breech during the baseline tests, the water-gas shift constant is shown in Figure E-3. With respect to these data, two things are noteworthy. First of all, the ratio H₂O/CO₂ remains relatively constant so that in the breech area the ratio CO/H₂ is expected to remain constant as well. The large amounts of water and carbon dioxide present act like a flywheel, so that large changes in the CO and H₂ concentrations are not at all seen as similar changes in water or CO₂. For refuse, the ratio H₂O/CO₂ is equal to 1.5 so that the water-gas shift constant could be represented by the equation:

$$K_{wg} = 1.5 \text{ CO/H}_2.$$

The fact that the calculated value of K_{wg} is even close to the theoretical value is a very encouraging support of the water-gas equilibrium theory. The jet data were not used for this because of the difficulty detecting H₂ with the modified probe.

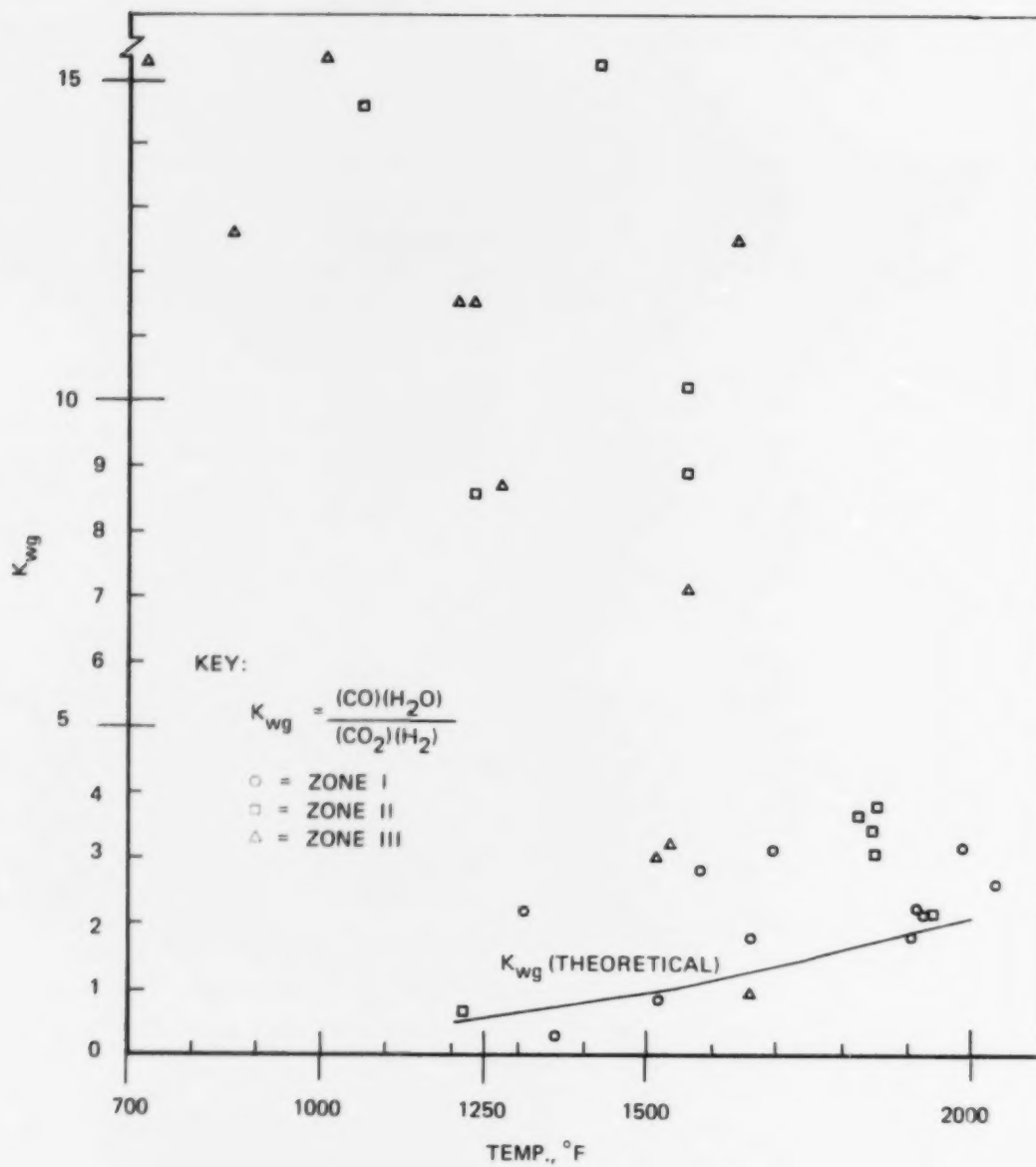


Figure E-3 Water - gas shift constant from breech data

Additional confirmation of the equilibrium can be seen from data taken directly over the drying grate. In these tests in which the concentration of CO and H_2 is sufficiently high to obtain good analytical measurements, the experimental value for K_{wg} is remarkably close to the theoretical value. A summary of the results of these tests was given in Table E-1. It is also interesting to note that the ratio of hydrogen to carbon is consistent with the H_2O/CO_2 ratio of 1.5.

The condition of uniformity is also easy to substantiate by considering a plot of H_2O/CO_2 as shown in Figure E-2. If, as the earlier models proposed, the drying zone and the pyrolysis zone were distinct from one another, then one would expect to see a reduction in the ratio of H_2O/CO_2 as the refuse travelled through the incinerator. For example, refuse on the first grate would have only a small amount of combustion or pyrolysis and a large amount of drying so that the ratio of H_2O/CO_2 would be larger than normal, reflecting the large amounts of water resulting from the drying and small amounts of CO_2 from combustion. On the other hand, the gases over the burnout grate should be completely dried and pyrolyzed so that very small amounts of hydrogen would be expected and the ratio of H_2O/CO_2 would be correspondingly small. In contradiction to this, the data indicate that there is no trend in H_2O/CO_2 , but rather that the ratio tends to be constant at approximately 1.5. The 1.5 value is consistent with the ratio of hydrogen to carbon for a refuse with 25% moisture (by weight). This tendency was observed both in the breech and over the bed and cannot, therefore, be attributed to overall mixing before the sample was taken.

Contrary to expectations, the H_2O/CO_2 ratio tends to be higher over the third grate than it does over the other two. We do not attribute this to any characteristic of the refuse-burning process, but rather to the fact that the hot ash dropping into the water-quench hopper evaporates a sufficient amount of water to cause a shift in the H_2O/CO_2 ratio over the third grate, an effect that would be witnessed in zone 3 because of the stratification of temperature layers in the breech of the furnace. Indeed, a material balance about the third grate indicates that the water introduced by this evaporation mechanism would be sufficiently large for such an effect to be observed.

DETERMINATION OF OXYGEN FRACTION

Significant amounts of oxygen were present in most of the samples taken above the fuel bed. Since, at furnace temperatures, oxygen does not coexist with supra-equilibrium concentrations of CO and hydrogen for more than a few milliseconds, the detection of both oxygen and fuel components in the samples is indicative of intermittent sampling from fuel-rich and oxygen-rich eddies. The oxygen observed above the bed probably originates in the underfire air which passes unreacted through blowholes or overfire air carried down by buoyancy forces. Table E-3 presents estimates of the amount of oxygen present in the gas samples expressed both as a fraction of the stoichiometric air requirements and a fraction of

the total air in the sample (determined from a nitrogen balance). With the exception of one sample, the available oxygen exceeded the stoichiometric oxygen requirements. The role of overfire air jets for the present conditions therefore should be principally to provide mixing, and caution should be exercised that the jets do not quench the combustion reactions.

Table E-3. AVAILABLE OXYGEN IN SAMPLES WITHDRAWN FROM ABOVE THE FIRST GRATE

Bed Test 1 Probe Position	Available O ₂	Unreacted O ₂	Unreacted O ₂
	O ₂ Required to Complete Combustion	O ₂ Supplied	Total Offgas
1	2.82	0.58	0.64
2	74	0.89	0.90
3	86	0.69	0.70
			0.75
Bed Test 2 Probe Position			
1	0.875	0.31	0.29
2	9.1	0.69	0.68
3	8.2	0.41	0.41
			0.46

A peculiarity in the data resulting from the design of the incinerator merits comment. The unreacted fraction in Table E-3 peaks at position 2, which was directly below one of the roof air jets. An attempt made to close the damper to the roof jets between bed tests 1 and 2 resulted in a reduction of the unreacted air. This tended to support the hypothesis that unreacted overfire air was entrained in the sample. Because of the presence of overfire air, it is not possible to determine, on the basis of these measurements alone, the fraction of the underfire air which is unreacted. The bed-burning data suggests that most of the unreacted oxygen originates in the overfire jets.

DEGREE OF MIXING

The data given in Table E-1 could be used to calculate an initial unmixedness factor according to the procedure outlined in Appendix B. In fact, the value for Φ_o has already been given in Table E-1. The corresponding values for $(\sqrt{c'^2})_o$ are shown in Table E-4. Values given for position 3, near the end of the first grate, are similar to the value for $(\sqrt{c'^2})_o$ calculated in Table E-3. One might be tempted to rationalize, explaining that position 3 is closest to the center of the furnace, particularly to the first burning, and is therefore more representative of the bulk of combustion gases found in zone I of the breech. The data from zone I were used to calculate the values given in Table E-3. Unfortunately, the data are too scattered and inconsistent for such a rationalization; no conclusions are apparent from the data.

Table E-4. INITIAL UNMIXEDNESS FACTORS

<u>Bed 1</u>		<u>CO/CO₂</u>		
	$\sqrt{C'_0}/\Delta C$			
Position		0.25	0.50	1.0
1	2.3	0.811	0.611	0.365
2	0.60	0.462	0.409	0.340
3	0.58	0.300	0.249	0.185
Average		0.524	0.423	0.297
 <u>Bed 2</u>				
Position				
1	- 2.0	1.273	2.982	4.951
2	1.14	0.556	0.457	0.332
3	1.18	0.286	0.184	0.061
		0.705	1.208	1.781

APPENDIX F

DERIVATION OF THE UNMIXEDNESS FACTOR

The following derivation is a summary of the development of the "unmixedness factor" concept based upon Reference . The variable, Φ , the ratio of oxygen in the sample to the oxygen required to complete combustion of the sample, can be calculated as follows:

If y is the mole fraction of offgas in the offgas/air mixture, then the mole fraction of air is $(1 - y)$. Also, since $4.76 \cdot n_{O_2}$ moles of air react with 1 mole of offgas, then the amount of offgas required to react with $(1 - y)$ moles of air is $[(1 - y)] / (4.76 \cdot n_{O_2})$. The remaining offgas after the available oxygen has been consumed is equal to the initial moles (y) minus the moles consumed, or in other words:

$$\begin{aligned} \text{Excess offgas} &= \left[y - \frac{(1 - y)}{4.76 \cdot n_{O_2}} \right] & (F-1) \\ &= \left(\frac{1 + 4.76 \cdot n_{O_2}}{4.76 \cdot n_{O_2}} \right) \left[y - \left(\frac{1}{1 + 4.76 \cdot n_{O_2}} \right) \right]. \end{aligned}$$

Since

$$y_s = \left[\frac{1}{1 + 4.76 \cdot n_{O_2}} \right] \quad (F-2)$$

$$\text{excess offgas} = \alpha (y - y_s) \quad (F-3)$$

where

$$\alpha = \text{constant } (1 + 4.76 \cdot n_{O_2}) / (4.76 \cdot n_{O_2}). \quad (F-4)$$

The oxygen required to complete combustion is n_{O_2} times the amount of offgas, so that:

$$O_t \text{ req} = (C - C_s)$$

Likewise, in a sample where y is less than y_s , the air concentration is still $(1 - y) \cdot C_t$, but in this case oxygen remains unreacted, i.e., the limiting reactant is offgas. In this case, $y \cdot C_t$ moles of offgas require $(4.76 \cdot n_{O_2}) \cdot (y \cdot C_t)$ moles of air. The remaining air is:

$$\begin{aligned} \text{Excess air} &= (1 - y) \cdot C_t - (4.76 \cdot n_{O_2}) \cdot (y \cdot C_t) \\ &= (1 + 4.76 \cdot n_{O_2}) (y - y_s) \cdot C_t & (F-5) \end{aligned}$$

and

$$[\%O_2]_{xs} = \left[\frac{1 + 4.76 \cdot n_{O_2}}{4.76} \right] \cdot (C - C_s) = \alpha' (C - C_s) \quad (F-6)$$

Note that Eq. (F-6) is exactly the same as Eq. (F-4), except that it applies to the region where $y > y_s$, while Eq. (F-4) applies to the region where $y < y_s$. Since the fluctuations about $\bar{C}$ are randomly distributed, the two equations could each be integrated over the entire range of fluctuations to determine the available oxygen in the sample and also the oxygen required to complete combustion. Hawthorne* et al showed, however, that it was more useful to calculate the ratio of these two. Since the oxygen concentration in the sample is zero whenever y is greater than y_s , then the available oxygen in the sample is determined from the integration of Eq. (F-4) between the concentration limits from zero to C_s . Likewise, the oxygen required to complete combustion for the entire sample is therefore determined from the integration of Eq. (F-6) between the limits from C_s to C_t .

Hence, the ratio of oxygen in the sample to oxygen required reduces to:

$$\Phi = - \frac{\int_0^{C_s} (C - C_s) f(c') dc'}{\int_{C_s}^{C_t} (C - C_s) f(c') dc'} \quad (F-7)$$

where $f(c')$ is the frequency distribution of c' .

The integrals can be evaluated if the form of the frequency distribution is assumed. In this regard, Hawthorne* et al cited several examples of the gaussian nature of turbulence. They therefore adopted the distribution:

$$f(c') = \frac{h}{\sqrt{\pi}} e^{-h^2 c'^2} \quad (F-8)$$

Then, applying this distribution, Φ upon integration becomes:

$$\Phi = \frac{-1 + \sqrt{\pi n} e^{-n^2} + \text{erf}(n)}{1 + e^{-n^2} + \text{erf}(n)} ; n = \frac{1}{\sqrt{2}} \frac{C}{\sqrt{c'^2}} \quad (F-9)$$

A plot of Φ vs $\frac{\sqrt{c'^2}}{\Delta C}$ is given as Figure 3 in Section V.

APPENDIX G

CONVERSION FACTORS FOR SI UNITS

Length

$$1 \text{ ft.} = 0.3048 \text{ m}$$

$$1 \text{ in.} = 0.0254$$

Volume

$$1 \text{ cu. ft.} = 0.02832 \text{ m}^3$$

Mass

$$1 \text{ lb} = 0.4536 \text{ kg}$$

$$1 \text{ lb/ft}^3 = 16.017 \text{ kg/m}^3$$

$$1 \text{ lb/hr} = 0.126 \times 10^{-3} \text{ kg/sec}$$

$$1 \text{ lb/hr-ft}^2 = 1.355 \times 10^{-3} \text{ kg/m}^2\text{-sec}$$

$$1 \text{ ton/day} = 0.0105 \text{ kg/sec}$$

$$1 \text{ ton/hr} = 0.438 \times 10^{-3} \text{ kg/sec}$$

Flow Rate

$$1 \text{ cfm} = 0.472 \times 10^{-3} \text{ m}^3/\text{sec}$$

$$1 \text{ ft/hr} = 0.0847 \times 10^{-3} \text{ m/sec}$$

$$1 \text{ ft/sec} = 0.3048 \text{ m/sec}$$

Energy

$$1 \text{ Btu} = 0.252 \text{ kcal}$$

$$1 \text{ Btu/lb} = 0.555 \text{ kcal/kg}$$

$$1 \text{ Btu/lb}^\circ\text{F} = 1.0 \text{ kcal/kg } ^\circ\text{C}$$

$$1 \text{ Btu/hr} = 0.070 \times 10^{-3} \text{ kcal/sec}$$

$$1 \text{ Btu/hr ft}^2 = 0.754 \times 10^{-3} \text{ kcal/m}^2 \text{ sec}$$

$$1 \text{ Btu/hr ft}^3 = 2.47 \times 10^{-3} \text{ kcal/m}^3 \text{ sec}$$

Temperature

$$1^\circ\text{F} = 0.555^\circ\text{C}$$

$$\text{Temp } (^\circ\text{F}) = 0.555 \text{ Temp } (^\circ\text{C}) + 32^\circ\text{F}$$

$$\text{or } 1.8 [\text{Temp } (^\circ\text{F}) - 32^\circ\text{F}] = \text{Temp } (^\circ\text{C})$$